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COLOR REMOVAL FROM KRAFT PULPING EFFLUENT BY LIME ADDITION

Charles L. Davis

Interstate Paper Corporation

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Color Removal From Kraft Pulping Effluent By Lime Addition

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Recovery of calcium used was carried out under mill conditions on a continuous basis following a statistically designed program. Results and full size design factors are given.

Performance of natural biochemical stabilization following lime treatment is shown graphically. Overall BOD_5 reduction for the tertiary treatment system is 98% with a final discharge average BOD_5 of 6 ppm.

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by

Interstate Paper Corporation
Riceboro, Georgia 31323

for the

Environmental Protection Agency

Program #12040 ENC
Grant #WPRD 183-01-68
December 1, 1971

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This report has been reviewed by the Environmental Protection Agency and approved for publication. Approval does not signify that the contents necessarily reflect the views and policies of the Environmental Protection Agency nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

ABSTRACT

A prototype color removal system was designed, constructed and operated as an integral part of a tertiary treatment system for total process effluent from a kraft linerboard mill. The basic system includes a lime precipitation process for the removal of color combined with primary clarification followed by natural biochemical lake stabilization and mechanical aeration.

Operating results show that the color removal system can operate successfully under widely varying conditions to give a relatively constant effluent color in the range of 125 ppm APHA color units at treatment levels of 1000 (± 50) ppm of calcium hydroxide with untreated effluent colors in the range of 1200 (± 200) ppm. Treatment at this level reduces lime cost to $53.73 per million gallons with lime at $15.35/ton (90% CaO). Performance is directly related to control of lime feed. Equipment evaluation indicated substantial savings in capital cost for future installations.

Recovery of calcium used was carried out under mill conditions on a continuous basis following a statistically designed program. Results and full size design factors are given.

Performance of natural biochemical stabilization following lime treatment is shown graphically. Overall BOD_5 reduction for the tertiary treatment system is 98% with a final discharge average BOD_5 of 6 ppm.

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SECTION I

CONCLUSION

A color removal system in a tertiary treatment process for a kraft pulp mill can operate successfully under widely varying conditions to give a relatively constant effluent color in the range of 125 ppm APHA color units at treatment levels of 1,000 ppm (± 50 ppm) of calcium hydroxide with untreated effluent colors in the range of 1,200 ppm (± 200 ppm). Treatment at this level offers advantages in cost savings and in obtaining sludges which are more suitable for dewatering. Lime costs are $53.73 per million gallons with lime at $15.35/ton (90% CaO). Performance is directly related to control of lime feed.

Operating experience and studies of the process have shown that flash mixing of lime slurry with the mill effluent prior to flocculation has no effect on the reduction of color. This is particularly so as flocculation has to be at stirring rates sufficient to keep carbonate mud and other solids in the mill effluent in suspension. These other solids including wood fiber aid clarification. Because of this, clarifiers for this service can be conservatively designed at rise rates of 0.5 gallons per minute per square foot of clear rise area. Flocculation time can be reduced to thirty-five minutes. The smaller equipment requirements offer substantial savings in future installations. However, cost should not be a factor in the selection of untreated effluent pumps. As calcium precipitation and stabilization following clarification of treated effluent is not part of the color removal process at Interstate, 140 acre lake surface area is required for six mgd for natural absorption of carbon dioxide from the atmosphere before natural biochemical treatment is effective.

Additional advantages of the color removal process are that it preconditions the effluent for biochemical treatment allowing rapid degradation. Foaming is eliminated and phosphorus concentration in the effluent is reduced thereby preventing possible eutrophication. High alkalinity in the process provides sterilization so treated domestic sewage does not require chlorination when handled through the system.

Because lime precipitation for color removal in tertiary treatment of kraft pulp mill effluent is a new development, the facilities at Interstate Paper Corporation's mill have been sponsored by the Environmental Protection Agency as a demonstration plant. In

keeping with this intent, operating information has been made available to interested parties and to others in the pulp and paper industry. Unpublished interim reports have been presented at an Auburn University engineering short course and at a regional meeting of the National Council for Air and Stream Improvement. The plant has been open for inspection. The number of visitors to date number 238 including paper mill representatives from across the United States and representatives from Canada, Finland, Sweden, and France. With the interest shown in the operating experiences at Interstate there will be further improvements and acceptance of the process.

The impact of the treated effluents from the mill on the environment is particularly noteworthy. The stabilization lake has become a natural preserve for egrets, ducks, and geese. There are also otters, alligators, and deer which feed in the area. The lake has been stocked with bream and bass by the Georgia State Game and Fish Commission. It is, to the author's knowledge, the first time that an effluent treatment facility is also being used as a sport fishing lake.

The extensive combination primary, secondary and tertiary treatment plant at Interstate Paper Corporation has successfully demonstrated that the pulp and paper industry can be productive and protect a healthy environment.

In designing a carbonation system for recovery of calcium carbonate and the stabilization of the highly alkaline kraft pulp mill effluent following color removal, the major factor is the formation of the calcium carbonate precipitate in the carbonation zone. With careful attention to this part of the physical and chemical reaction process, maximum calcium recovery can be obtained without the use of coagulant aids at high clarification flow rates. There are several standard designs of water treatment equipment on the market by different companies which can be used with some modification.

The following general specifications for process equipment are conservative at a 10 million gallons per day rating.

Carbonation Zone

Counter current gas and water flow
Carbonation depth - 10.5 ft
Residence time - 16-20 min

Recycle to Carbonator Feed

Volume - 25% of total flow
Solids - 135-275 lbs/1000 gal

Clarification Zone

Rise rate - 2.50 gal/min/sq ft
Residence time - 60 min
Sludge zone - 20% of total volume
Blow down - 0.33% of total feed
Sludge concentration - 2.25 lbs/gal

If a combined carbonation and clarification unit is selected, the approximate size of the carbonation zone would be 40 ft in diameter 10.5 ft deep in a 72 ft diameter clarifier with a 15 ft water depth. For this size clarifier a maximum rake torque rating of 100,000 ft pounds will be more than adequate.

The residence time for the carbonation zone is based only on the physical requirements of the gas diffuser.

With an average mill lime kiln stack gas temperature of 180° in a range of 160-200° F, the average calcium loss will be 1.025 lbs $CaCO_3$ per 1,000 gallons of waste treated (123 ppm). By withdrawing the stack gas before it enters the venturi scrubber and removing the particulate matter in a separate cyclone, gas temperature could be raised to 325° F. This would decrease the calcium loss to approximately 0.43 lbs $CaCO_3$/1,000 gal or 50 ppm. However, there will be added cost for blowers, piping and diffusers. Polyelectrolites could possibly reduce calcium loss. In this case operating cost would be increased. The economics of the individual mills will dictate which plan to follow.

SECTION II

INTRODUCTION

Color of kraft pulp mill effluents has been considered objectionable in receiving streams mainly from an esthetic standpoint - particularly in recreational areas. In downstream sources of municipal and industrial water supplies, treatment cost may be increased. Color could interfere with biological activity by retarding transmission of sun light into the stream and indicate undesirable dissolved organic chemicals which could possibly be harmful to commercial and sport fishing.

It was this last factor which was the greatest concern of the Georgia State Water Quality Control Board as the effluent from a kraft pulp mill in the Riceboro vicinity would enter a virgin coastal area, well known as a highly productive breeding ground of commercial shrimp, oysters, and crabs, as well as for its excellent sport fishing. To assure protection for the aquatic life, chemical treatment to remove color was made a requirement in the construction permit for the Interstate Paper Corporation kraft linerboard mill at Riceboro, Georgia.

The color removal process using slaked lime which was developed, designed and constructed to meet the Georgia State requirements had industry wide application provided it was economically feasible. Because of this the Federal Water Quality Administration, now EPA-WQO, entered into a research study with Interstate Paper Corporation in June of 1968.

This report is a summarization of the operation of the color removal process for a two year period. It goes beyond the scope of the Federal Water Quality Administration study to give operating results of natural biochemical stabilization following the color removal process for the second year of operation. A third part of the report covers a joint pilot plant scale research project undertaken by the Interstate Paper Corporation and the Continental Can Company, Inc. under their respective FWQA grants to investigate the recovery of calcium from lime treated kraft pulp mill effluent under mill operating conditions.

In the color removal process being used at the Interstate mill, calcium hydroxide in a slurry of constant concentration is mixed with the total process effluent in direct proportion to flow. The

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mixture is retained in a flocculator for thirty-five minutes and then clarified in a center feed clarifier. Color of the combined effluents average 1160 APHA units with a range of 260 to 2300 units. The colored substances (mainly lignin by-products) are precipitated as calcium salts and are removed from the system in the underflow of the clarifier together with fiber and other settleable solids and pumped to a holding pond. The decolorized effluent overflowing the clarifier is saturated with calcium hydroxide at a pH of 12.2. On entering an oxidation lake, following the treatment process, the treated effluent absorbs carbon dioxide from the atmosphere and the calcium is precipitated as carbonate, while the pH drops to an acceptable range for biochemical stabilization to occur in the lower lake areas. Following stabilization, the effluent is aerated by a single floating aerator on discharge under controlled conditions to the receiving stream.

SECTION III

OBJECTIVES

The objectives of the project were to develop, install, and demonstrate a new chemical process for removing color from kraft pulp mill effluent.

The color removal process was designed as one phase of a multi-step effluent treatment system which includes primary clarification and secondary biochemical stabilization along with post reaeration. Application of the process was for a new 400 ton kraft linerboard mill to meet the following limitations stipulated by the Georgia State Water Quality Control Board:

Effluent discharge	10 mgd
Biochemical oxygen demand	800 lbs/day
Suspended solids	10 ppm
Color (APHA cobalt units)	30 ppm

At the time this effluent treatment system went into service, March 15, 1968, there was no known proven color removal process in operation for kraft pulp mill effluents in the United States. In the public interest, it was the intent to demonstrate that the color removal process would:

a) Remove undesirable and possible harmful color matter from kraft mill effluent.

b) Preserve the esthetic qualities of the stream, thus enhancing its value for recreational purposes while at the same time permitting use of the stream for vital industrial purposes.

c) Be a technically and economically feasible color removal process applicable throughout the kraft pulp industry.

d) Provide an early test of the efficiency of the process in actual use.

e) Be adaptable to treatment systems of other mills in operation.

The objectives of the carbonation pilot plant study were to obtain the design criteria for the most efficient system at minimum cost for the recovery of calcium used in the process and the conditioning of the

effluent from the color removal process for immediate biochemical stabilization, either by the natural or by the accelerated processes.

SECTION IV

CHRONOLOGY

Mill Site

Interstate Container Corporation is a producer of corrugated boxes in the northeastern states. It was with the intent to supply its own requirements for linerboard that Interstate Paper Corporation was formed and the mill at Riceboro, Georgia, was built and put into operation March 15, 1968.

The mill is located on a 1,900 acre site in Liberty County, sixteen miles inland from the Atlantic Coast and borders Riceboro Creek and the North Newport River. It is thirty-two miles south of Savannah. See vicinity map Figure 1 page 10.

The North Newport River flows through marsh lands to the east and empties into St. Catherine's Sound. No other industrial waste enters the coastal waters between the Savannah River to the north and the Altamaha River to the south, a distance of fifty miles.

The marine biologists and conservationists refer to these marsh areas and coastal waters as the most productive breeding grounds of commerical shrimp, oysters, and crabs in the world. There is good sport fishing from Half Moon Landing on the North Newport River, 15.1 stream miles below the mill site into the sound and in the off-shore waters.

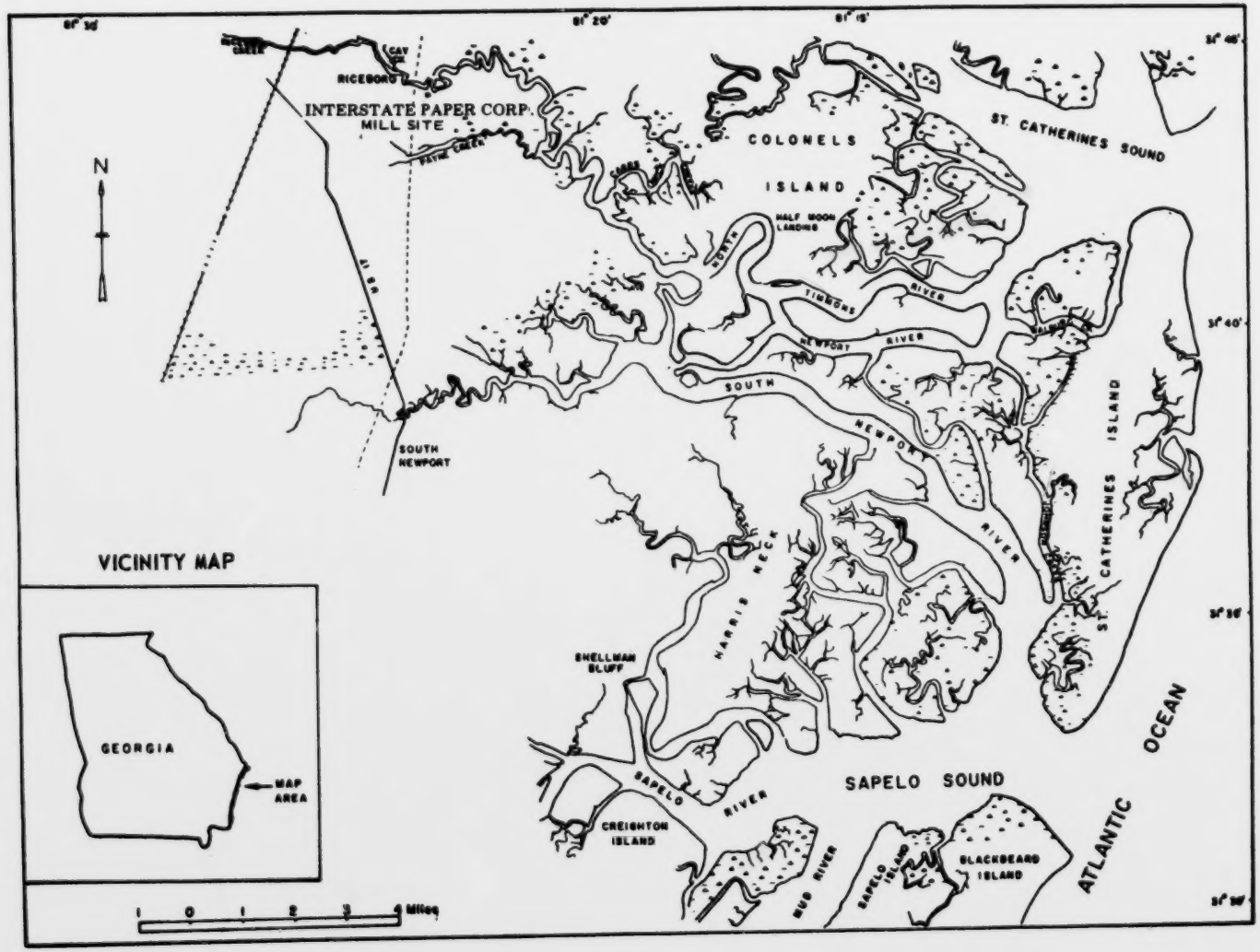
Stream Surveys and Cooperative Research Studies

As a major part of the mill site evaluation, stream surveys were conducted to establish existing chemical, biological and physical conditions and to estimate stream potential for effluent assimilation. Six foot tides and complex estuarine conditions made this a difficult assignment. Two surveys can be summarized in a few words.

1. For all practical purposes, Riceboro Creek is a dead end of the estuary. Discharge of 10 mgd mill waste would, at times, occupy the full cross section of the stream, requiring fifteen to twenty days to reach the Sound.

2. Sea water does not extend far upstream from Half Moon Landing; and for this reason, tides could not be relied upon

FIGURE 1


INTERSTATE PAPER CORP.
MILL SITE
RICEBORO
COLONELS
ISLAND
ST. CATHERINES SOUND
HALF MOON LANDING
TIMBERS
RIVER
NEWPORT RIVER
SOUTH NEWPORT
SOUTH NEWPORT RIVER
ST. CATHERINES ISLAND
HARRIS NECK
SHELLMAN BLUFF
SAPELO RIVER
CREIGHTON ISLAND
MUD RIVER
SAPELO SOUND
SAPELO ISLAND
BLACKBEARD ISLAND
ATLANTIC OCEAN
N
VICINITY MAP
GEORGIA
MAP AREA
1 0 1 2 3 4 Miles

to introduce fresh sea water to supply dissolved oxygen and dilute mill effluent. Oxygen replenishment of the stream is mainly through surface contact with the atmosphere.

3. The river water dissolved oxygen (DO) of 4.8 to 5.5 ppm and the biochemical oxygen demand (BOD) of 0.5 to 1.0 ppm indicated no significant source of pollution other than general marsh drainage.

4. To maintain a minimum DO of 2 ppm, three to four miles below the mill and a minimum of 4 ppm DO above the Half Moon Landing would limit the effluent discharge to between 500 and 1,000 lbs of BOD per day. It would also be required that the waste have a DO concentration of 6 ppm.

5. Treated mill waste could be discharged into the North Newport River without serious effects on the water below Half Moon Landing or damage to fishing near the ocean. Above Half Moon Landing there is no extensive sport fishing and no competitive uses of the river water. Below this point there is extensive sport fishing and some oyster harvesting.

The complexity of estuarine conditions required extensive study beyond the preliminary surveys. For this reason and to record any changes in the receiving water, cooperative investigative and research studies of the North Newport and St. Catherine's Sound were established, one by the Georgia State Water Quality Control Board and the University of Georgia Marine Institute at Sapelo Island, and one by the Georgia State Water Control Board and the U. S. Geological Survey. Both of these studies were for four years; two of these years were prior to mill discharge for comparison with the next two years after mill startup. In the first program, water samples have been collected and analyzed, fish and marine organisms have been collected, counted and identified for population density and bioasseys have been performed to determine tolerance levels of aquatic life to unbleached kraft mill waste. The first part of the biological studies was presented at the TAPPI Air & Water Conference, Jacksonville, Florida, in April, 1969 (1)(2)(3). The second cooperative program is mainly a hydrological study of the estuary tidal movement and fresh water entry. Seven stations have been established on the North Newport River and in St. Catherine's Sound, five to record stage and two to record flow direction and velocity. Two of these stations automatically monitor DO, pH, turbidity, temperature and conductivity. Wind direction and velocity are also recorded. Water quality and hydrological data

can be obtained from the U. S. Geological Survey office in Atlanta, Georgia.

Construction Permit

Based on the preliminary surveys, primary clarification, and natural biochemical stabilization followed by mechanical aeration were submitted for effluent treatment in the application for construction permit from the Georgia State Water Control Board. In meetings and discussions following, the Board took the position that effluent treatment as proposed would possibly not be adequate protection for the estuarine shrimp and fish breeding grounds. This was their greatest concern. As an added precaution, two suggestions were made; one, provide a minimum of 200 to 1 dilution at Half Moon Landing by diverting flow from the Medway River on the north; or, two, provide chemical treatment to give a residual waste color of 30 ppm APHA units.

On evaluating the two proposals, Interstate selected chemical treatment and the Board set the following limitations:

Waste discharge not to exceed 10 mgd
Effluent BOD not to exceed 800 lbs/day
Suspended solids not to exceed 10 ppm
Color not to exceed 30 ppm APHA color units

The intent of the limitation on effluent color was to define the degree of chemical treatment for the removal of tannic acid and lignin compounds which are not readily biodegradeable.

The color standard is particularly noteworthy as there was no background information on the effectiveness of color removal, and it is the first time that it has been used as a measure of chemical treatment required by a regulatory agency. In the past, effluent color has been considered a detriment to steam photosynethesis, or to the stream's esthetic value.

Mill Description

The mill was designed and built by The Rust Engineering Company of Pittsburgh, Pennsylvania, for a rated capacity of 400 tons of unbleached kraft linerboard per day. A detailed description of equipment has been published in several trade journals [4][5]. Besides the modern design, an important feature is the steps taken to reduce

fresh water usage and losses. Process water requirements, exclusive of cooling water, were 12,500 gallons per ton compared to 25,000 gallons per ton average for similar production in the industry. Biochemical oxygen demand to the waste treatment plant was anticipated to be in the order of 25 lbs per ton compared to the 35 lbs per ton average.

At the end of the operating period March 15, 1970, present production capacity has increased to 563 tons per day without additions to the facilities. Process water has decreased to 9,800 gallons per ton with 29 lbs BOD per ton to the effluent treatment plant.

Effluent Treatment Facilities

The color removal process was developed and designed from bench test data without the benefit of pilot plant studies. Time was a major factor. Subsequent research at the Paper Institute in Appleton, Wisconsin under a FWQA grant [6] and at Syracuse University, New York, funded by the National Council for Stream and Air Improvement of the Pulp and Paper Industry has developed the mechanics of the lime reactions with kraft effluent color bodies. This information will lead to further refinements of the process and the reduction of cost.

Under the original concept for effluent treatment as first proposed, flow was by gravity to a remotely placed primary clarifier discharging to a natural stabilization lake. To provide for color removal, the primary clarifier was relocated to within the mill yard. A lift station, lime handling system, flash mixing and flocculator were added ahead of the clarifier. The design was based on a maximum waste flow of 10 mgd and a BOD_5 loading of 14,000 lbs per day. Description of the process and equipment as installed has been given in an interim progress report to the TAPPI Air and Water Conference in Jacksonville, Florida, April 1969 [7] and to the AIChE 63rd Annual Meeting in Chicago, Illinois 1970 [8].

In review, the equipment as installed and in the order of flow includes:

1. Pumping station - consisting of a sump with a fixed bar screen to which all process wastes flow by gravity, three 3,500 gpm vertical lift pumps and recirculation piping for level control.

2. A lime feed and flocculation system including a 210 ton lime silo, screw conveyors, 65 ton/day slaker, 750 gallon slurry tank with agitator, inline mixer, 45 ft diameter, 33 ft high floc mix tank with agitator, piping and instrumentation to control the

lime feed in proportion to waste flow.

3. Centerfeed primary clarifier - 200 ft in diameter and 15 ft water depth providing for twelve hours retention at 5 mgd average flow and six hours retention at maximum flow of 10 mgd.

4. Sludge lagoon with decant pumps - 167 acre feet capacity.

5. Process effluent holding lagoon with barge mounted pump and return piping to process sump; 48 million gallon capacity.

6. Natural stabilization lake - 650 acres with an average depth of 4.4 feet providing 180 days retention at 5 mgd average flow not including rainfall or losses due to evaporation.

7. Post mechanical aeration - including a 75 HP mechanical floating aerator and a concrete lined retention basin forming integral part of the discharge structure.

8. Discharge structure - consisting of a fixed bar screen, motorized gate valve, flow nozzle and recorder, motorized flow control valve, continuous sampling device and 0.5 mile of 36" pipe for submerged discharge into Riceboro Creek.

Estimated Cost and Federal Participation

Total capital investment for effluent treatment at the Riceboro mill is in the order of $2,500,000 representing 10% of the total mill cost. Construction cost of the color removal plant was estimated at $454,100 including the plans, specifications and construction supervision. Site cost and preparations were estimated at $27,500, giving a total capital cost of $481,600. Operations were estimated at $269,000 per year for chemicals, power, and labor. Other costs including administration, contingencies, and post construction studies and reports were estimated at $133,100, bringing the first year cost to $883,700. Assuming a straight line ten year depreciation and based on 140,000 tons per year, the total cost of the lime color removal system would be $3.19 per ton of production for the first year.

Since the lime color removal system was unique and of industry wide application, it was believed that the Riceboro mill system could well serve as a demonstration plant. Accordingly, a formal application for a Research and Development Grant was submitted to the Federal Water Pollution Control Administration on August 11, 1967. On

June 13, 1968 a grant of $466,895 was offered to and accepted by Interstate Paper.

SECTION V

LIME TREATMENT - MATERIALS AND METHODS

A. Systems

Mill Operating Conditions

The production of paper is a dynamic performance of many men controlling machines and using large quantities of raw materials The chemical and physical properties of effluent from this process are ever changing as the position of a bouncing ball. Variations in effluent flow at the Riceboro mill are as much as 100 percent from the average because of machine wire and clothing changes and plugging of the effluent sump bar screen. The latter causes short duration surges. Intermittent flows of green liquor dregs and lime mud from the recausticizing area and stock from stock chest cleanouts and floor washups in the pulp mill and machine areas can cause inaccurate flow measurements. Effluent chemical properties of the combined effluent stream are far from constant because of the variation in volume from all sources and because of pullovers from the evaporators following boilouts and washups in the chemical recovery area.

All process effluents flow by gravity to a sump and lift station as well as rain run-off from paved and curbed areas around chemical and liquor storage areas. Sewage septic tanks also drain to the process sewers. Uncontaminated cooling water is diverted to Riceboro Creek. As a precaution against spills or storage tank ruptures, there is a motorized gate operated by a conductivity probe at the outfall of the cooling water. Should contamination occur, this gate closes and an alarm is sounded so that the water can be diverted to a holding pond.

Figure 2 page 18 is a flow diagram of the tertiary treatment system including all changes to date. There may be further modifications as experience dictates.

Process Effluent Flow

As a quick rundown of the treatment system, the process effluent stream flows through a fixed bar screen at the entrance to the sump. Collected trash is removed manually as needed. Flow is then through the lift pumps and under pressure through

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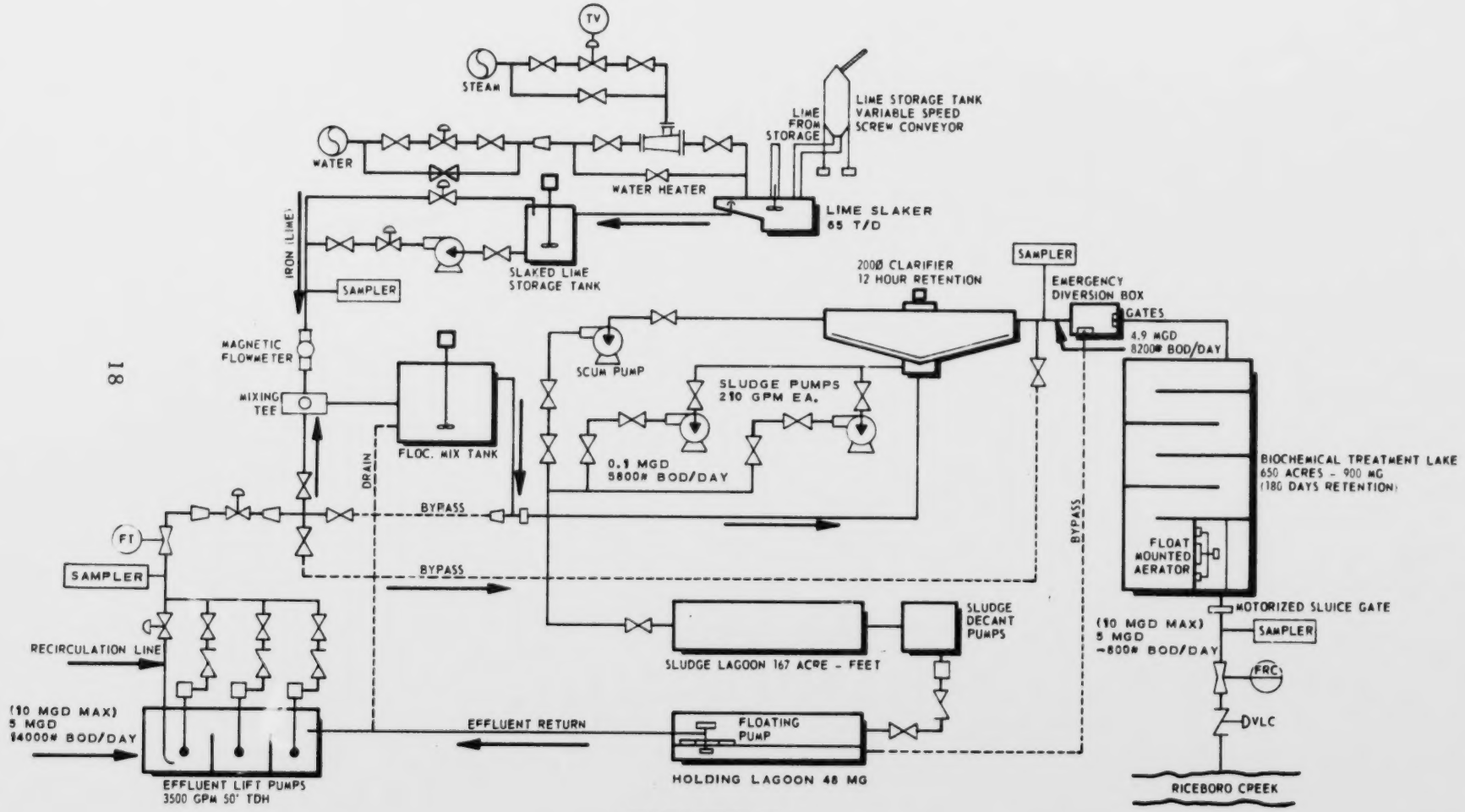

FIGURE 2

EFFLUENT TREATMENT FLOW DIAGRAM

an eccentric orifice and mixing tee into the floc tank. Slaked lime slurry is fed into the effluent stream at the mixing tee. Calcium hydroxide in the slurry combines with effluent color constituents and forms a brown precipitate in the floc tank. The flow is then by gravity through the clarifier and on through the remaining part of the system.

In the clarifier the settled floc, inert solids and fiber are moved to the center by a mechanical rake and withdrawn by two Moyno pumps discharging to one of the holding lagoons. Foam and scum are removed from the surface of the clarifier by a skimmer attached to the sludge rakes and are likewise discharged to the same lagoon. Supernatant from the sludge is pumped to the adjacent process waste holding lagoon. The decolorized and clarified effluent overflows the clarifier through the peripheral collection launderer and pipe line to the stabilization lake or it is diverted to the second holding lagoon if the color is not acceptable. Untreated effluent is also diverted to this lagoon when there is an equipment outage or a major mill upset. Diverted flows are pumped back to the sump for retreatment. Normally, by-pass lines are not shown on a simplified flow diagram. With treatment for color removal, this is an important feature when a color standard is to be met. It will be noted on referring to Figure 2 that flows can be diverted from any point in the system.

The decolorized effluent entering the stabilization lake at a pH of 12.2 has in solution 30 to 72% of the calcium hydroxide added to the system depending on the treatment level. The calcium is ultimately precipitated as calcium carbonate by carbon dioxide absorbed from the atmosphere. Calcium precipitation occurs in the first 100 acre section of the lake. The effluent is then suitable for biochemical stabilization which occurs in the adjacent areas of the lake. Before controlled discharge to Riceboro Creek, mechanical aeration brings the dissolved oxygen up to saturation.

Chemical Flow

In the order of chemical flow, calcined lime (CaO), is withdrawn from the silo and fed to the slaker by a variable screw conveyor controlled by an air bubbler type specific gravity measurement device on the discharge side of the slaker. Water flow to the slaker is controlled by a d/p cell and level controller in the slaked lime slurry tank. Slaking temperature is

maintained at 200° F. by a direct steam water heater controlled by a controller and thermocouple placed in the slaking compartment of the slaker. Under normal operating conditions the lime slurry to the slurry tank is held between 1.05 and 1.06 grams calcium hydroxide per liter at 190° F.

Water is added to the slurry tank to trim out the lime concentration at 1.04 grams per liter at 170° F. Slurry concentration at this point in the system is also controlled by a similar gravity measurement device.

Slurry is withdrawn from the slurry tank by a 120 gpm constant discharge pump at 180 ft head. On the discharge side of this pump, the lime slurry flow is split by two control valves. One flow is through a magnetic flow meter to the mixing tee ahead of the floc mix tank. The other flow is recirculated to the slurry tank and to a timer operated sampler. These two flows are controlled by a controller and the eccentric orifice in the process waste line on the discharge side of the lift station. Lime slurry flow is in direct proportion to waste flow. All variables are recorded. Should the lime slurry demand fall below 20 gpm, the slurry automatically goes into full recycle, the slurry sampler and flow integrator trip out and the pipe line to the floc tank is flushed with water. When the demand increases to above 20 gpm, the flushout system reverses itself.

Instrument settings for automatic control are made from charts of calcium hydroxide slurry concentrations from 1.03 to 1.07 grams per liter at temperatures of 170 and 190° F; and tables of effluent and lime slurry flow for treatment levels of 1,000 to 2,600 ppm calcium hydroxide. With interruption of automatic control for maintenance or repair, the system is put on manual control based on average waste flows and charts of lime screw output.

The system is monitored by weighing grab samples of lime slurry from the slaker and slurry tank for comparison with instrument readings. At twenty-four hour intervals, operating charts are collected, a lime slurry feed composite sample is checked for calcium hydroxide concentration, and integrator readings of lime slurry and process effluent flows are recorded.

The transition from a concept of a process to a full scale operating plant can be a rugged but challenging experience.

This is particularly so when the system is a prototype design which must go on stream with a new mill startup. The Riceboro mill treating plant was no exception. The following is a commentary on the components of the system.

B. Operations and Results

Initial Operation

As the process was a prototype design, it was necessary to determine the limitations and adequacy of the equipment as it has been installed. To do this, range of lime feed, color of untreated and treated process waste, and waste flow were recorded during the first 90 days of operation.

There are two continuous timer-operated sampling stations on the process waste flow set to collect fifty gallons composite samples for each 24 hour period. The one for the total waste flow to the lime treating process is of a splitter box flow design to minimize plugging from suspended solids. The one for the treated waste overflow from the clarifier is of the solenoid valve design.

A third continuous timer operated sampling station was added at the lime slurry feed tank. It is a splitter box design similar to that in use on the untreated waste flow line. This sample station is used to monitor the calcium hydroxide concentration of the lime feed to the floc tank.

On a routine basis, samples were collected from the two process waste flow sampling stations every twenty-four hours and tested for pH, alkalinity, and color. Twenty-four hour composite samples of lime slurry were analyzed for calcium hydroxide concentration. Waste flow and lime feed were recorded for the corresponding period of sampling.

On every eight hour shift and as often as indicated by mill operating conditions, grab samples of treated and untreated waste were tested for color to check level of treatment.

On as needed basis or when time permitted the following tests were performed:

1. Jar coagulation test to check level of treatment.
2. Available CaO of calcined lime.

3. BOD_5 of untreated and treated waste.
4. Dichromate oxygen consumed (DOC).

Operating Period - June 27, 1968 through September 30, 1968

Untreated waste color averaged 1080 APHA platinum cobalt units with a high and low of 8,000 and 200. The average was slightly above the indicated normal operating range of 600 to 1,000 ppm.

Treated waste color averaged 95 color units adjusted to pH 7.6 with a low and high of 450 and 40 units. Percent reduction was fairly constant in the range of 90%, irrespective of the untreated waste color.

Average color reductions per 100 ppm of calcium hydroxide feed for the ranges of color and lime feed rates are shown on page 102. Lime usage as 90% CaO averaged 37.51 tons per operating day.

Process waste flow to the color removal system averaged 5.15 mg for a total of 453.32 mg for the period. The process waste was 53.3% of the combined process and cooling water flow. During the period, 23.7 mg of treated waste was diverted to the holding pond and 42.1 mg was returned from the holding ponds for treatment giving a net reduction in retained waste of 18.3 mg.

From operating experience through this period, the following limitations were observed:

1. Maximum feed of lime as $Ca(OH)_2$ resulted in a concentration of 2,000 ppm at a waste flow of 6 MGD. Maximum feed was increased to give a concentration of 2,600 ppm by installing a parallel lime feed line.

2. Above 6 MGD, waste flow was the governing factor for lime concentration.

3. Below 4 MGD, lime feed was erratic. Density of lime in slaker and slurry tank increased beyond the desired level.

4. Rapid changes in waste flow gave a quick response from

the dilution control valve between the lime slurry tank and transfer pump suction. As a result dilution water backed up into the slurry tank; the desnity of lime in the slurry tank decreased and the level rose; water to the slaker shut off and the lime screw continued to feed lime at the rate of 0.6 tons per hour.

5. Lime feed to the slaker fluctuated with the level of lime in the silo, lime particle size and percent available lime as CaO.

6. Lime feed to the inline mixer decreased with scaling in the transfer line. As a result treatment was interrupted to clean out the line.

7. The system had an excessive inherent lag.

Based on the above observations, the lime feed control system was redesigned. The most important features were:

1. Dilution water was to be added to the slurry tank instead of to the transfer pump suction.

2. A density controller and recorder was added to vary the dilution water flow and record lime slurry density.

3. Lime feed to the inline mixer was to be controlled by an air operated valve in direct ratio to waste flow.

4. Lime slurry was to be circulated back to the slurry tank to provide for continuous sampling.

5. A magnetic flow meter was to be installed to record lime slurry feed.

Equipment was ordered and delivery was expected to be completed by October 11, 1968. Installation was scheduled for completion by November 8, 1968.

Operating Period October 1, 1968 through October 31, 1968

Modifications to Equipment

During the equipment startup, the floc agitator was modified in

that one set of blades was removed when the floc tank discharge was lowered to stream-line flow to the clarifier. Agitation was reduced by forty-five percent and the tank volume was reduced twenty-five percent. To compensate for the loss in agitation, the agitator drive-speed was set at the maximum rpm.

When the floc tank was drained for the mill shutdown, October 13, excessive buildup of solids was observed. The drain line was closed by sludge and the tank had to be emptied through the inlet line and by-pass to the holding pond. To reduce buildup of solids, it will be necessary to replace the upper blades. This will require the removal and machining of the agitator shaft to provide a minimum of 48 inches spacing above the lower set of blades. Work was scheduled for completion on November 22, 1968.

The clarifier had settled because of compaction of the bearing soil. Most of this settling in the past had been at the center column. A check of the launder levels on October 7 showed that the north half was 1.3 to 1.4 inches lower than the south half The launder ring was scheduled to be re-leveled by December 7, 1968.

Operating Results:

During the month of October, the mill was shut down from the 13th through the 20th because of a labor strike. For this reason test data covers twenty-four (24) operating days.

The untreated process waste color averaged 1380 APHA platinum cobalt units with a high of 5,000 and a low of 400. The average color is 28% higher than the preceding period reflecting the higher losses during mill shutdown and startup. Treated process waste color averaged 120 color units with a high of 300 and a low of 50. Percent reduction in color was in the order of 90-92%.

Average color reduction per 100 ppm of calcium hydroxide concentration for the range of color and lime concentrations is shown in the Appendix page 103. Lime usage as 90% CaO averaged 47.32 T/D compared to 37.51 T/D average for the preceding period. The higher lime concentration figures are questionable as they give a lime usage higher than the end of the month inventory withdrawals. It is expected that closer correlation will be attained when the new lime feed metering system is completed.

Average process waste flow for the period was 4.74 mgd. This is 49.4 percent of the total mill flow including uncontaminated cooling water discharged to Riceboro Creek. Of the total process waste, 22.43 mg was diverted to the holding ponds and 14.93 mg was returned from the holding ponds for re-treatment giving an increase of 7.5 mg retained.

Operating Period November 1, 1968 through November 30, 1968

Modifications to Equipment:

1. Lime Feed System

   The modifications to the lime feed system were completed except for the installation of one control valve and the controller for the magnetic flow meter. Shipment was delayed on both of these pieces of equipment.

2. Inline Mixer

   Impeller of the inline mixer was cracked when the baffles were removed during the start-up period to reduce pressure drop. This impeller was replaced on November 21, 1968 when the treatment plant was taken out of service to replace bearings on the lime screw conveyor and to straighten the lime screw.

3. Floc Agitator

   While the treatment plant was out of service for repairs, the upper agitator blades were replaced. Work was completed one day ahead of schedule.

4. Clarifier Sludge Lines

   The sludge line discharge was extended 150 ft on November 15, 1968 to move ahead of the deposited sludge in the east holding pond.

Operating Results:

Some of the typical mill operating problems occurred during the month of November. As a result, process waste to the color removal plant had a color exceeding 1,000 ppm six operating days in the reporting period. Average process waste

color averaged 1006 APH platinum cobalt units with a high of 6,000 and a low of 250. Treated process waste color averaged 104 color units with a high of 200 and a low of 50. The high of 200 color units is the cut off point for diversion to the holding ponds.

Average color reduction per 100 ppm of calcium hydroxide concentration for the range of color and lime concentrations are shown in Appendix, page 104. Lime usage averaged 37.25 T/D.

The effect of lime treatment for color removal on biochemical oxygen demand (BOD_5) is shown in the following table with corresponding color units. From this data there appears to be no correlation between BOD and color. Additional data may show some relationship.

TABLE 1

BOD REDUCTION VS COLOR REDUCTION

PROCESS WASTE

	UNTREATED		TREATED		% REDUCTION	
Date 1968	Color APHA	BOD ppm	Color APHA	BOD ppm	Color	BOD
11/6	700	245	80	205	89	15
11/12	900	310	150	260	83	17
11/13	800	310	100	235	88	25
11/14	3000	335	200	230	93	31
11/19	1500	325	75	235	95	27
11/20	600	280	75	200	88	27
11/26	700	335	100	270	86	20
Av.	1060	305	110	235	88	23
Av. lbs. BOD/Day		17,100		13,100		

Average process waste flow for the period was 6.6 mgd. This is 72.3 percent of the total mill flow including uncontaminated cooling water discharged to Riceboro Creek. Of the total process waste 17.85 mg was diverted to the holding ponds and 27.19 mg was returned from the holding ponds for retreatment, giving a reduction of 9.34 mg in the amount retained.

Operating Period - December 1 through December 31, 1968

Modifications to Equipment:

1. Lime Feed System

   Delivery of materials delayed the completion of the system well into the middle of December. Because of this and with the pending Christmas mill shutdown and holidays, final completion was deferred until January 9, 1969.

2. Clarifier Launder

   Leveling of the clarifier launder requires outage time on the waste treatment plant. To minimize flow of process waste to the holding ponds, this work was rescheduled for January 9, 1969 to coincide with the tie in of new instrumentation of the lime feed modifications.

Operating Results:

During the month of December, color concentrations were the highest on record and the percent color reduction the lowest on record. Of the twenty-seven and one half operating days, color of the untreated process waste exceeded 1,000 APHA platinum cobalt units for fourteen days. Average for the period was 1,200 with a high of 2,750 and a low of 500. process waste color averaged 160 units with a high of 250 and a low of 75. Average percent color reduction was 86.3.

Average color reduction per 100 ppm calcium hydroxide concentration for the range of color and lime concentrations is shown in the Appendix, page 105. Lime usage averaged 42.9 T/D as 90% calcium oxide (CaO).

The temporary laboratory was destroyed by fire on December 5, 1968. For this reason, there was no testing done other than color determinations. One set of samples of untreated and treated process waste was sent to Shuey and Company, Inc., Analytical and Consulting Chemists in Savannah, Georgia, for BOD tests. Results are shown on Table 2, page 28, with corresponding color concentrations.

TABLE 2

BOD REDUCTION VS COLOR REDUCTION

PROCESS WASTE

Date 1968	UNTREATED Color APHA	UNTREATED BOD ppm	TREATED Color APHA	TREATED BOD ppm	% REDUCTION Color	% REDUCTION BOD
12/22	1500	300	150	310	90.0	neg.

Average process waste flow for the period was 6.26 mgd. This is 71.5 precent of the total mill flow including uncontaminated cooling water discharged to Riceboro Creek. Of the total process waste 14.0 mg was diverted to the holding ponds and 26.4 mg was returned from the holding ponds for re-treatment giving a reduction of 12.4 mg in the amount retained.

Operating Period - January 1 through January 31, 1969

Modification of Equipment:

The new instrumentation to control lime slurry feed by density was completed on January 9. At the same time, a new lime conveyor screw was installed to replace one which had been bent by pieces of broken lime kiln chain.

While the treating plant was down, the clarifier launder was leveled and cleaned. Short circuiting due to unbalance has been eliminated. There is still some intermittent short circuiting due to windage. With the changes, clarifier performance has been materially enchanced.

Operating Results:

From January 1 through January 5 the color removal system was operated with the lime feed controls as originally designed and installed. The treating plant went down at 10:00 A. M., January 6 to tie in the new instrumentation for lime density control, sampling and recording lime slurry flow. Treatment was resumed at 7:00 A. M., January 7. Following is a brief presentation of operating results before and after changes to the lime

feed instrumentation.

Mill conditions for January 1 through January 5 were normal with average untreated process waste color of 820 APHA platinum cobalt units. There were no unusual operating incidents. Treated process waste color averaged 94 units with an average lime usage of 36.4 T/D as 90% CaO. Average process waste flow was 7.39 mgd including 1.01 mgd returned from the holding ponds for re-treatment.

During the time interval of January 8 through January 31, the color removal plant was shut down twice for unscheduled mill shutdowns and seven times for the following repairs and modifications to equipment.

1. Broken lime conveyor screw.
2. Replacement of variable speed drive bearing on lime screw conveyor.
3. Replacement of slaker agitator motor.
4. Installation of process waste flow meter flush line.
5. Tie-in of piping to supply evaporator condensate for slaking lime and diluting lime slurry.

The last mentioned change was made to reduce scaling of the lime feed system.

Service was also interrupted on thirteen other days. The major causes were restricted condensate supply, plugging of process waste flow meter pitot tubes and build-up of lime density in slaker. Corrective steps were taken to reduce outages for these reasons.

When the color removal system was in service, lime feed rates and lime density were varied to calibrate instruments and to check ranges of treatment. On three consecutive days, January 27, 28, and 29, controls were set to maintain a concentration of 1800 ppm of $Ca(OH)_2$ in the effluent. An average for the three days was 1840 ppm with a variation of 40 ppm. Untreated process waste colors for this run were 1280, 2100, and 1780 APHA cobalt units. Treated waste colors in the same order were 220, 180 and 200 units representing an average 87.6% color reduction.

Average color reduction per 100 ppm calcium hydroxide for the

range of color and lime concentrations is shown in the Appendix, page 106 for the reporting period. Lime usage averaged 35.8 T/D as 90% calcium oxide (CaO).

Laboratory test work was limited during January because repairs to equipment damaged in the fire December 8 had not been completed and chemical reagents had not been completely restocked.

Available tests for BOD and COD are shown below with corresponding color concentrations.

TABLE 3

COD & BCD REDUCTICN VS COLOR REDUCTION
PROCESS WASTE

Date	UNTREATED			TREATED			% REDUCTION		
	Color APHA Units	CCD ppm	BCD ppm	Color APHA Units	COD ppm	BOD ppm	Color	CCD	BCD
1/1/69	1000	909	360	125	—	220	87.5		38.9
1/17/69	3600	1440	456	150	496	280	95.8	51.7	38.6
1/29/69	1780	1296	420	200	574	300	88.7	55.7	28.6

Average process waste flow for the reporting period was 6.49 mgd. This is 73.4 percent of the total mill flow including uncontaminated cooling water discharged to Riceboro Creek. Of the total process waste 27.1 mg was returned from the holding ponds for treatment giving an increase of 2.4 mg in the amount retained. Holding ponds were at their maximum capacity on January 13.

Operating Period - February 1 through February 28, 1969

Modifications to Equipment:

With the instrumentation for lime feed control installed in January, hidden operating problems were found and isolated. The major one was that there was no control of the lime slurry specific gravity in the slaker with the existing instrumentation. Ratios of water to lime were not a straight line function. As a result, lime slurry specific gravity increased to the point where the slaker would overload,and the dilution water to the slurry tank could not maintain the desired specific gravity. A difference of 0.01 in lime slurry specific gravity can give a

swing of 250 ppm in $Ca(OH)_2$ concentration in the treated waste. To correct for this, specific gravity controls were added to the slaker, making the lime screw rpm a slave of specific gravity instead of to the flow of water to the slaker. With these further modifications it was necessary to re-verify instrument readings of specific gravity against weighed samples of lime slurry.

Operating Results:

February was a short operating month with a seven day mill shutdown for maintenance repairs. The lime color removal system was operated for one additional day to treat wash up waste water. During the reporting periods untreated waste colors averaged 2010 APHA platinum cobalt units. This is 1000 units higher than the normal operating range. Treated process waste color averaged 169 units with an average lime usage of 37.9 T/D as 90% CaO. Average treated process waste flow for 22 operating days was 7.10 mgd including waste returned from the holding ponds for re-treatment. Average process waste flow during the same period was 6.56 mgd. This is 68% of the total mill waste flow including uncontaminated cooling water discharged to Riceboro Creek. Of the total waste flow for the 28 day period, 19.58 mg were diverted to the holding ponds and 21.39 mg were returned for re-treatment giving a net increase in holding pond capacity of 1.81 mg.

Average color reduction per 100 ppm calcium hydroxide for the range of color and lime concentration is shown in the Appendix, page 107, for the reporting period. The wide range of lime concentrations reflects the lack of lime feed control and illustrates the effect of the slaker specific gravity on the system.

Part of the laboratory test work included specific gravity determinations and calcium analysis of the lime slurry feed for calibration purposes. Available tests for BOD and COD are shown on page 32 with corresponding color concentrations.

TABLE 4

COD & BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date	UNTREATED			TREATED			% REDUCTION		
	Color APHA Units	COD ppm	BOD ppm	Color APHA Units	COD ppm	BOD ppm	Color	COD	BOD
2/4/69	2000	1825	421	200	682	323	90.0	62.6	23.0
2/5/69	2000	1634		300	851		85.0	47.6	
2/6/69	2250	1944	663	175	832	330	92.2	57.0	50.3
2/11/69	1200	961	283	125	685	238	89.6	28.7	15.9
2/13/69	2500	1515		250	923		90.0	39.1	
2/25/69	2000	985	404	100	462	246	95.0	53.0	39.0
2/26/69	1000	1464	339	80	223	259	92.0	84.7	24.9
2/27/79	6800	1496	546	125	401	244	98.0	73.0	58.0

Operating Period - March 1 through March 31, 1969

Modifications to Equipment:

Three changes were made to the lime feed system during the reporting period. These were:

1. Pressure diaphram used for specific gravity measurement of the slaked lime in the slurry tank was replaced with an air bubbler tube in a flow through pot mounted on the outside of the tank.

2. Slurry tank level control was rearranged as originally installed.

3. Electric vibrator was installed on lime silo.

The first change was made to reduce outage time. The bubbler tube and flow through pot is the same as used for control of specific gravity of the slaked lime in the slaker. They are so designed that they can be cleaned and adjusted quickly while the feed system is on manual control. Until this change was made the slurry tank had to be drained to remove the pressure diaphram for cleaning. Total time for cleaning and adjusting was one and one-half to two hours.

The second change was to revert back to a wider range for slurry tank level control. The third change is an addition to level out the flow of lime from the silo to the slaker.

Operating Results:

During March the color removal system was out of service for one hour and twenty minutes to replace a diaphram with an air bubbler system for the slurry tank specific gravity control. During the reporting period, untreated process waste colors averaged 1150 APHA platinum cobalt units. This is a 42.8% reduction compared to the average untreated waste color for the preceding period. Treated process waste color averaged 116 units with an average lime usage of 32.8 T/D as 90% CaO. Total process waste treated was 195.1 mg including 30.8 mg returned from the holding ponds for re-treatment. During the period, 2.7 mg was diverted to the holding ponds giving a net increase in available holding capacity of 28.1 mg. Average process waste flow from the mill averaged 5.38 mgd which is 66.8% of the total mill waste flow including uncontaminated cooling water diverted to Riceboro Creek.

On March 2 pulp mill operations changed the pulp washing sequence. As a result the average total process waste flow was reduced by 1.2 mgd. There was a decrease of approximately 2.7 mgd in the pulp mill sewer. This decrease was offset by an increased usage in the paper mill. Normally with all systems operating, there would be no effect on the performance of the color removal system. With the paper machine down for a wire change the process waste flow dropped between 2.0 and 2.5 mgd. In this range of waste flow, the lime feed system was below the lime feed flow range. This was compensated to maintain lime feed by changing the ratio setting of slaked lime flow to waste flow. However, in doing so, the system was taken off the desired constant feed rate.

To meet the changes in waste flows, it would be necessary to change the sprockets on the variable screw drive and to drop the trip out signal on the automatic back-flush system. The pulp washing sequence was on a trial basis and would be continued until April 13. A third washing sequence would be tried and run through May 11, 1969.

Average color reduction per 100 ppm calcium hydroxide for

the range of color and lime concentration is shown in the Appendix, page 108, for the reporting period.

Laboratory test work included jar tests to check treatment results. Specific gravity of slaked lime samples collected over a 24 hour period were also determined to verify instrument read-outs. Tests for BOD and COD are shown below with corresponding color concentration.

TABLE 5

COD & BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	COD ppm	BOD ppm	TREATED Color APHA Units	COD ppm	BOD ppm	% REDUCTION Color	COD	BOD
3/4	1020	1018	411	100	592	202	90.2	41.8	26.5
3/5	920	800	304	80	776	228	91.3	30.0	25.0
3/6	1660	1443	382	30	—	115	98.2	—	69.9
3/11	1540	1217	332	130	590	223	91.6	51.5	32.8
3/12	980	1023	246	130	589	223	86.7	42.4	9.3
3/13	1080	1018	280	90	—	204	91.7	—	27.1
3/18	1140	1019	314	80	496	226	93.0	51.3	28.0
3/19	460	817	196	90	602	151	80.4	26.3	23.0
3/20	960	1050	212	N.S.	N.S.	—	—	—	—
3/25	1860	1333	351	100	483	230	94.6	63.8	34.5
3/26	530	820	296	90	530	198	83.0	35.4	33.1
3/27	2640	1976	408	240	—	246	90.9	—	39.7
Av.	1233	1128	311	105	582	204	90.9	42.8	31.7

Comparing the above data with corresponding data for the preceding period and taking into account decrease in waste flows, total BOD was decreased by 73.8%.

Operating Period - April 1, through April 30, 1969

Operating Results:

During the month of April the waste treatment plant was programmed to treat the process waste with 1500 ppm calcium hydroxide $Ca(OH)_2$ from April 1 through April 16 and with 1200 ppm calcium hydroxide from April 17 through April 30. There was a day by day variation. However, for the two periods during the month, the average for the first period was 1450 ppm and 1246 for the second period. In comparing the two

programmed periods, a higher percent reduction in color was obtained at the lower lime concentration of 1246 ppm. Lime usage decreased 0.58 tons CaO (90%) per million gallons of process waste treated. This indicates an optimum level of treatment beyond which additional lime is wasted.

On an overall operation for the month of April, the untreated process waste color averaged 1084 APHA platinum cobalt units which, is slightly lower than the average untreated waste for the preceding period. Treated process waste color averaged 97 units with an average lime usage of 29.68 T/D as 90% CaO. Total process waste was 189.27 mg including 11.84 mg returned from the holding ponds for retreatment. During the period, 7.73 mg was diverted to the holding ponds giving a net increase in available holding capacity of 4.11 mg. Average process waste flow from the mill averaged 6.16 mgd which is 73.1% of the total mill waste flow including uncontaminated cooling water diverted to Riceboro Creek.

Down time of the chemical coagulation plant in April was fifteen hours and twenty minutes. Ten hours and fifty minutes were for three periods of spillage of chemicals and kraft liquor. During this time, the process waste was diverted to the holding ponds to level out the load on the waste treatment system.

Four hours and thirty minutes were required to clean the inline mixer, to change the sprocket drive on the lime screw and for a piping change.

Even taking the above into consideration, performance of the waste treatment plant had improved materially during the reporting period. Average color reduction per 100 ppm calcium hydroxide for the range of color and lime concentration is shown on page 109.

Laboratory test work on lime slurry specific gravity and calcium analyses was routine to check lime feed instrument settings. Tests for BOD and COD are shown on the following page with corresponding color concentrations.

TABLE 6

COD & BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	UNTREATED COD ppm	UNTREATED BOD ppm	TREATED Color APHA Units	TREATED COD ppm	TREATED BOD ppm	% REDUCTION Color	% REDUCTION COD	% REDUCTION BOD
4/8	1440	1134	278	180	561	246	87.5	50.5	11.5
4/9	460	524	212	140	—	—	69.5	—	—
4/15	4200	1807	478	160	755	266	96.2	58.2	44.4
4/16	220	688	274	100	305	267	54.5	55.7	2.6
4/17	700	498	305	80	—	294	86.6	—	3.6
4/22	960	840	304	90	359	224	90.6	57.3	26.3
4 23	1140	1105	292	60	330	202	94.7	70.1	30.8
4/24	1320	1074	367	100	—	209	92.4	—	43.1
Av.	1305	959	314	114	462	244	84.0	38.4	23.2

Operating Period - May 1 through May 31, 1969

Operating Results:

During the month of May, the waste treatment plant was programmed to treat the process waste with 1200 ppm $Ca(OH)_2$.

All during the month, trouble was experienced in holding the calibration of the specific gravity controls on the slaker and slaked lime slurry tank. Air lines to and from the D. P. cells were replaced when it was found that acid, used to clean the bubbler tubes of the specific gravity controls, had backed up into these lines. It was not until the 30th of the month that it was found that acid had also backed up into the D. P. cells. On inspection both diaphrams were leaking through numerous cracks and pin holes caused by the acid.

Another problem in holding set specific gravity occurred during the first part of the month when the water supply to the slaker and slurry tank intermittently shut off. Water for the lime feed system was pumped from a surge tank which was also used as a source of water for the caustic area. High water usage in this area would drain down the surge tank to a point where there was not sufficient water for the lime feed system. This was corrected by resetting the make-up water control valves.

As a result of the above operating difficulties the average

concentration of lime to the color removal system was 1418 ppm $Ca(OH)_2$. Variations in treatment were large, ranging from 870 to 2160 ppm. Considering the volume of waste treated, the increase of 1200 to 1418 ppm represents an excess of 112 tons of lime (90% CaO). Process waste flow was reduced from 189.27 mg for the month of April to 159.98 mg for the month of May. As a net result, lime usage dropped from 29.69 T/D to 24.13 T/D as 90% CaO for May. This drop in process waste flow was due to a trial run on the mill pulp washing sequence.

Down time of the chemical treatment plant in April was 39 hours; of this, $19\frac{1}{2}$ hours were for work on the inline mixer and cleaning the lime silo discharge gate. The lime feed system was shut down $19\frac{1}{2}$ hours while strong waste was diverted to the holding ponds to level out the load on the color removal system. Waste, both treated and untreated, was diverted for a total of 62 hours during May. The strong waste was from a green liquor clarifier overflow and from washouts of the recovery boiler cyclone and smelt dissolving tanks.

Other than the overage on lime feed, performance for May was good. Untreated process waste color, exclusive of waste diverted, averaged 736 APHA platinum cobalt units. Treated process waste color averaged 86 units. Average color reduction per 100 ppm calcium hydroxide for the range of color and lime concentration is shown in the Appendix, page 110. The total process waste of 159.98 mg treated included 14.43 mg returned from the holding ponds for retreatment. During the period 12.26 mg was diverted to the holding ponds giving a net increase of 2.17 mg in available holding capacity. Process waste flow from the mill averaged 5.11 mgd which is 59.7% of the total mill waste flow including uncontaminated cooling water diverted to Riceboro Creek.

Laboratory test work on lime slurry specific gravity and calcium analysis were routine to check lime feed instrument settings. Tests for COD and BOD are shown on page 38 with corresponding color concentrations.

TABLE 7

COD & BCD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	UNTREATED COD ppm	UNTREATED BOD ppm	TREATED Color APHA Units	TREATED COD ppm	TREATED BOD ppm	% REDUCTION Color	% REDUCTION COD	% REDUCTION BOD
5/6	1060	816	271	100	413	169	90.6	47.1	59.5
5/7	820	867	233	120	578	213	85.3	33.3	9.0
5/8	1100	955	288	120	NT	251	89.1	——	12.8
5/13	620	916	225	50	371	200	91.9	59.5	11.0
5/14	1140	1143	212	80	443	NT	93.0	61.2	——
5/15	840	847	NT	90	NT	NT	87.8	——	——
5/20	180	667	NT	50	356	193	72.2	46.6	——
5/21	160	1039	235	60	494	161	62.5	52.5	31.5
5/27	600	709	298	75	470	228	87.5	33.7	23.4
5/28	380	568	221	80	NT	194	78.9	——	12.2
Av.	690	853	248	83	446	201	83.9	47.7	22.8

NT - No test

Inline Mixer Evaluation:

Design of the color removal system was based on laboratory data without verification by pilot plant studies. Laboratory work clearly indicated that precipitation of color bodies would not occur without rapid mixing prior to flocculation. The purpose of this evaluation was to determine the effect of mechanical agitation vs mixing by baffling.

May 12, 1969 was selected for the date to start the evaluation of the inline mixer. It was at this time that the pulp washing sequence in the pulp mill was scheduled to use the decker as a fourth stage washer. This trial run was to be for 28 days corresponding to a mill accounting period. The evaluation test period was set up to run 14 days with the inline mixer in service followed by 14 days without the inline mixer in service under comparable mill operating conditions. Prior to starting the tests, the inline mixer was removed, cleaned and reinstalled. At the end of the first 14 days, May 27, the inline mixer was removed and a blind flange was bolted in place to close the opening for the agitator. The housing with internal baffles was not removed because the baffles would prevent laminar flow at the point of lime slurry injection.

Comparing the first five days' performance without the inline mixer in service with the preceding two weeks with the mixer in service, no difference has been observed in the degree of treatment obtained.

Operating Period - June 1 through July 31, 1969

Operating Results:

During the two month periods there have been continuing problems with instrumentation of the lime feed system and the process flow metering system. Basically, except for failures of diaphrams of d/p cell transmitters, and control valve diaphrams, most of the problems stem from varying water and air pressures.

With variations in lime feed and malfunctions of the process waste flow transmitter, it has not been possible to maintain the desired treatment level of 1200 ppm concentration of calcium hydroxide. In June the average lime consumption was 4.81 tons 90% CaO per million gallons of waste treated or 1,387 ppm $Ca(OH)_2$. The average for July was 4.62 tons of lime or 1,332 ppm $Ca(OH)_2$. All during July the lime feed system had to be on partial manual or full manual control.

In June untreated process waste color averaged 1,235 APHA platinum cobalt units. Treated process waste color averaged 108 units. Average color reduction per 100 ppm calcium hydroxide for the range of color and lime concentration is shown on page 111. The total process waste of 208.533 mg treated included 29.20 mg returned from the holding ponds for re-treatment. During the period 8.60 mg was diverted to the holding ponds giving a net increase of 20.60 mg in available holding capacity. Process waste flow from the mill averaged 6.33 mgd which is 68.2% of the total mill flow including uncontaminated water diverted to Riceboro Creek.

In July untreated process waste color averaged 1388 APHA platinum cobalt units. Treated process waste color averaged 127 units. Average color reduction per 100 ppm calcium hydroxide for the range of color and lime concentration is shown on page 112. The total process waste of 169.94 mg included 23.83 mg returned from the holding ponds for re-treatment. During the period 6.67 mg was diverted to the holding ponds giving a net increase of 17.16 mg in available holding capacity.

Process waste flow from the mill averaged 5.88 mgd which is 61.9% of the total mill flow including uncontaminated water diverted to Riceboro Creek. Due to a mill shutdown for maintenance and repairs there were 26 operating days in July.

Comparing the above data for the two periods, process waste decreased 7.1% in July. As there was no change in mill process, some of the decrease can be attributed to errors in flow measurement. Average overall color reduction for the two periods was approximately the same at 91%. Percent reduction is a function of the untreated waste color. At the lower ranges of color the percent reduction is approximately 10% less than in the higher ranges. The percent reduction is shown for each range of color in the Appendix, page 111, and page 112. Tests for COD and BOD are shown on page 41 with corresponding color concentrations.

Lime Feed Evaluation:

The lime feed evaluation has not been concluded for the reason that control has not come up to expectations nor has it given constant and repeatable results. Problems in holding constant feed stem from fluctuations in water supply and pressure to the water control valves to the slaker and to the slurry tank, fluctuation in lime slurry flow to the specific gravity measuring pots, and variations in air pressure to the pneumatic controls.

It was thought that the water supply problem was resolved by resetting the make-up water control valve. This did not solve the problem. The make-up valve was changed from a 2" to a 4" to increase availability of mill supply water. A separate supply pump motor and pressure regulator were ordered. Installation was scheduled for the week of August 18.

To control lime slurry flow, a constant headbox was installed ahead of the specific gravity measuring pot of the lime slurry. This proved very satisfactory. A constant discharge pump was installed on the slaker to control lime slurry flow to the specific gravity measuring pot at this point in the system. The pump was not satisfactory and was to be replaced and modified. Completion was scheduled for the week of August 11. Air pressure regulators were in the process of being installed on the air supply to bubbler tubes of each specific gravity measuring pot.

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TABLE 8
COD & BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	COD ppm	BOD ppm	TREATED Color APHA Units	COD ppm	BOD ppm	% REDUCTION Color	COD	BOD
6/3	450	620	216	60	286	198	86.7	53.8	1.1
6/4	800	834	302	50	444	185	93.7	46.7	38.7
6/5	250	1111	212	75	NT	173	70.0	—	18.4
6/10	1250	1007	368	100	636	264	92.0	36.8	28.3
6/11	1000	917	328	100	500	227	90.0	45.4	30.8
6/12	2500	1608	333	150	NT	265	94.0	—	20.4
6/17	800	760	329	85	365	295	89.4	51.9	1.0
6/18	1500	989	398	75	502	178	95.0	49.2	55.2
6/19	900	719	213	75	NT	208	91.6	—	2.3
6/24	2500	1917	406	200	696	280	92.0	63.7	31.0
6/25	1600	1148	382	80	NT	329	95.0	—	13.9
Av.	1232	1057	317	95	490	236	89.9	49.6	21.9
7/1	850	844	288	80	386	212	90.6	54.3	26.4
7/2	1000	843	280	100	NT	208	90.0	—	25.7
7/9	550	549	183	75	368	183	86.4	32.9	0
7/10	1500	NT	292	80	NT	254	94.6	—	13.0
7/15	1000	1091	405	120	NT	221	88.0	—	45.4
7/16	1000	NT	281	100	395	233	90.0	—	17.1
7/17	700	555	263	95	NT	173	86.4	—	34.2
7/22	1600	1070	447	140	659	302	91.3	38.4	32.4
7/23	5000	2055	643	150	715	268	97.0	65.2	58.3
7/24	1000	992	288	125	NT	239	87.5	—	17.0
7/29	2000	1658	356	130	532	284	93.5	67.9	20.2
7/30	1200	1141	340	130	578	265	89.2	49.4	22.1
7/31	1250	1110	376	100	NT	281	92.0	—	25.3
Av.	1435	1083	342	110	519	240	90.5	51.4	25.9

NT - No test

Inline Mixer Evaluation:

It was hoped that as many variables as possible could be eliminated, one of which was the variations in the lime feed to the system. If this could have been done, comparisons of two operating periods, one with the inline mixer in service and one with the inline mixer out of service, would have been sufficient. This not being the case, two other periods of 28 days each, before and after the period selected, were compared. Results are shown in the following tabulation.

TABLE 9

INLINE MIXER EVALUATION
TREATMENT DATA

Operating Period	5/12 5/26	5/28 6/8	4/14 5/11	6/9 7/6
Inline Mixer Service	in	out	in	out
*Waste Color, Untreated	804	650	1113	1457
Treated	100	77	96	118
Treatment-(Average) $Ca(OH)_2$ ppm	1618	1507	1340	1277
Color Reduction ppm/100 ppm $Ca(OH)_2$	45.7	35.1	77.6	111.3
Color Reduction %	87.3	83.2	90.3	90.7

*APHA cobalt units ppm - average

Operating periods were selected to have corresponding mill operating conditions. Percent color reduction is less at low untreated waste colors as shown in the Appendix, page 111. For this reason the difference in percent reduction between the periods shown in the above first two columns could be expected with or without the inline mixer. At the higher untreated waste colors, as shown in the next two columns, the difference is less. This is also normal. Treatment above 1200 ppm $Ca(OH)_2$ gives relative lower reduction of color per 100 ppm $Ca(OH)_2$ used.

The data is not straightforward as it must be interpreted. The above comments can be summarized: treatment results were normal for the combination of untreated waste colors and treatment levels with the inline mixer in service and with it out of service. In short, the inline mixer does not contribute to the

degree of treatment obtained.

Floc Agitator Evaluation:

At the beginning of the program it was planned to vary the degree of agitation to determine the effect on floc formation and its effect on color removal. Inspections made during mill shutdowns indicate that this will not be practical. Full agitation is required to keep settleable solids in suspension. Even then there is a build-up of solids around the bottom of the tank at the outer edges.

Clarifier and Pipe Line Inspection:

During the mill shutdown July 4 through July 8, 1969, the clarifier was emptied. The concrete bottom was inspected for cracks and the rakes were checked for clearance. The bottom was found in good condition. One rake was dragging at the outer end. This was corrected by removing one inch from the outer blade.

On July 8, one opening was cut in the steel horizontal run of the overflow line from the floc tank to the clarifier center feed well, and two openings were cut in the 36" concrete discharge pipe line from the clarifier to the lake. The inspection holes in the concrete pipe line were 1/3 and 2/3 of the distance from the diversion structure to the outfall.

The interior surface of the steel pipe was evenly covered with approximately 1/8" layer of light colored scale probably of high calcium content. At the high water mark on the downward leg of the overflow pipe, the scale was approximately 1/2 inch.

In the concrete pipe at the inspection hole nearest the diversion structure, there was approximately 1/4" of hard solid scale and lumps or rings of a softer scale containing fiber up to 2" thick. At this point in the line the pipe flows full. At the second point of inspection of the concrete pipe line, the same scale and rings were noted but to a lesser degree of thickness. At this point the pipe line flows approximately 3/4" full. At the outfall, scale had formed up to the flow line or about 1/3 of the pipe cross section. It was less at this point than at the other points of observation.

When scale was removed from the concrete pipe, the exposed

surface was smooth and showed no signs of erosion. Measurements of the pipe diameter at the three points were 36-1/16, 36-1/8, and 36-1/8 inches in the direction of flow.

The pipe lines have been in service for fifteen months. Overall, there were no signs of erosion. The scale build-up at this time is not significant. Should the waste be stabilized by carbonation, the scale build-up will be materially reduced, if not eliminated.

Operating Period - August 1 through August 31, 1969

Modifications to Equipment:

Changes to the water supply system to the lime slaker and slurry tank were to include:

1. Replacement of 2" water makeup control valve with a 4" control valve.

2. New pump and motor.

3. Water pressure regulator.

The above equipment was received. The pump foundation and pump were installed. However, the tie-in to the water surge tank could not be made until the scheduled mill shutdown, September 1.

The constant discharge pump to control lime slurry flow to the specific gravity measuring pot on the slaker was rebuilt by the supplier and was reinstalled. This was not complete until the last of the month.

Full automatic control operation of the lime feed system is contingent on the completion of the water supply system.

Other changes made were:

1. Installation of a small booster pump and pressure tank to maintain a water supply at 30 psi to the purge system on the process waste flow measuring nozzle.

2. Installation of differential pressure regulators on the above water purge system.

In addition, the waste flow recorder was taken out of service for recalibration.

In the past there has been considerable difficulty in maintaining the waste flow measurement instrumentation. Changes in the system have been made to try to resolve these operating difficulties. To further check out the system, the flow nozzle was scheduled to be removed for inspection during the scheduled mill shutdown September 1.

Operating Results:

All during the month of August, lime was fed at a constant rate to give a concentration of 1200 ppm calcium hydroxide for a calculated average process waste flow of 6.6 mgd. To do this the lime screw conveyor rpm was adjusted to give a lime feed which would maintain a slaked lime slurry specific gravity of 1.05 with a constant slurry flow rate of 63 gpm.

The actual average concentration of calcium hydroxide maintained during the reporting period was 1305 ppm. For three days the concentration averaged 891 because of a high volume of surface water from heavy rains draining into the process sewer. Deleting these three days, the average for the period was 1349 ppm $Ca(OH)_2$. Average lime consumption was 4.59 tons 90% CaO per million gallons of waste treated. The average percent reduction in process waste color with this treatment was 91.3.

The above color reduction is on the same reporting basis that has been used since the start of the program. Color is expressed in APHA platinum cobalt units. The color of untreated waste samples are at the existing pH of the samples which for the period averaged 10.1. The color of the treated waste samples is at an adjusted pH of 7.6. Both of these pH values represent actual pH of waste as they are and will be observed. The reason for following this procedure is that some color reduction can possibly occur with the addition of acid causing precipitation of some of the color bodies. Since June 21, color of untreated waste has been determined at the unadjusted pH and at a pH of 7.6. The difference in color at the lower pH was 14.8% less than at the higher pH. With both colors, untreated and treated waste, at a pH of 7.6, the overall percent color reduction for the reporting period was 89.9 compared to 91.3 for the color reduction not adjusting the pH of the untreated

waste samples. For this period the color reduction per 100 ppm calcium hydroxide concentration is reported with the untreated and treated waste both adjusted to 7.6 and is shown for each range of color in the Appendix, page 113. It will be noted that the overall pattern of values has not changed, other than a small decrease in some instances.

The lime feed system was not put on full automatic control as consistent performance could not be obtained without first making changes to the water supply system to the lime slaker, and the slurry tank. Otherwise, the performance of the waste treatment plant was at an acceptable level. Total down time was $8\frac{1}{4}$ hours; $5\frac{1}{4}$ hours were for repairs of the lime screw conveyor variable speed drive, and 3 hours were for inspection of the lime slurry pump impeller. The lime screw impeller was badly worn and was scheduled for replacement. The waste treating plant was shut down at midnight of August 31, with the mill shutdown for scheduled repairs.

Total process waste treated was 193.78 mg including 29.55 mg of waste returned from the holding ponds. During the down time referred to above, 3.90 mg of waste were diverted to the holding ponds. Based on the waste in and out of the holding ponds, available holding capacity should have increased 25.65 mg. Actually, the ponds are approximately three-fourths full because of heavy rainfall during August. Process waste flow from the mill averaged 5.42 mg, which is 60.0% of the total mill flow including uncontaminated water diverted to Riceboro Creek.

Test data for COD and BOD are shown on page 47 with corresponding color concentrations.

TABLE 10

COD & BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	UNTREATED COD ppm	UNTREATED BOD ppm	TREATED Color APHA Units	TREATED COD ppm	TREATED BCD ppm	% REDUCTION Color	% REDUCTION COD	% REDUCTION BOD
8/5	1150	831	260	80	415	209	93.0	50.1	19.6
8/6	880	792	243	80	401	184	90.9	49.4	24.2
8/7	980	931	251	125	NT	210	87.2	———	16.3
8/12	1000	986	353	120	754	262	88.0	41.8	25.8
8/13	1400	1253	334	110	509	177	92.1	59.3	47.0
8/14	1400	1161	278	110	NT	202	92.1	———	27.3
8/19	1080	928	275	110	451	202	89.8	51.4	26.5
8/20	900	724	192	90	524	150	90.0	28.0	21.9
8/21	1220	1478	293	100	NT	180	91.8	———	38.6
8/26	1200	1156	376	130	NT	255	89.2	———	32.2
8/27	1500	NT	349	160	NT	259	89.3	———	25.8
8/28	1140	———	338	130	NT	259	88.6	———	23.4
AV.	1154	1024	295	112	479	212	90.2	46.7	27.4

Note NT - Heating shelf for COD test out for repairs.

Operating Periods - September 1 through October 31, 1969

Modifications to Equipment:

1. Water supply to lime feed syster...

   Piping, installation of pressure regulator, connection of pump suction to surge tank and change over from a 2 inch to a 4 inch make-up water valve were completed on October 22, 1969.

2. Specific gravity measurement of slaked lime from the slaker.

   Pumps used to sample lime slurry have not been satisfactory because of poor performance and excessive maintenance cost. A new centrifugal pump constructed of a metal equivalent to Carpenter 22 was purchased. This pump has a capacity of 10 gallons per minute with a 6 ft head at 1050 rpm. A constant headbox similar to the one used on the specific gravity measurement of the lime slurry from the slurry tank was to be used to reduce the flow to the specific gravity sensing device. Completion of this modification was

scheduled for November 5, 1969.

2. Process waste flow measurement:

The Badger-Penn "Lo Loss" flow tube was removed for inspection on September 18, 1969. Solid material, mostly lime mud, had deposited in the invert of the pipe and throat of the tube. The low pressure zone of the tube was completely filled with solids and the upstream taps were partially plugged. The buildup of solids was attributed to low water velocity.

The "Lo Loss" tube had been sized for a maximum flow of 15,000 gallons per minute with a differential pressure of 114.1 inches of water. Operating flows averaged 4,600 gpm and did not exceed 8,000 gpm. At these low flows the maximum differential pressure was 32.5 inches. The control valve to maintain level in the lift pump sump, located downstream from the flow tube, was affecting the differential pressure read out. Both of these factors decreased accuracy of flow measurements.

To correct the conditions noted, a major change was indicated. On October 28, 1969, the flow tube was removed and replaced with an eccentric orifice plate with vena contracta taps. Upstream flush outs were also provided. This orifice is designed for a range 0 to 8,000 gallons per minute with a maximum pressure differential of 100 inches of water. Performance appears to be satisfactory.

Operating Results:

Difficulties in maintaining the waste flow measurement continued after all known corrective measures had been made, including installation of a separate purge water system with pressure regulators for the flow nozzle. For automatic control of the lime feed system, the waste flow measuring devices must function properly. As this was questionable, the lime feed system was operated on manual control for an average flow of 6.6 mdg, continuing the operating procedure for the preceding month.

During September, the color removal system was down one day for the Labor Day mill shutdown plus another seventeen hours

for repairs. The repairs included work on the flocculator agitator, clean-out of floc tank, work on lime screw drive, and removing scale buildup in 3,000 ft of 36 inch concrete pipe line from the clarifier to the lake. The untreated waste flow nozzle was also taken out of service for inspection and cleaning. During these repairs, waste flow was diverted to the holding ponds. There was an additional nineteen hours and fifteen minutes of diverting to the holding ponds due to the high color of the treated waste due to upsets in the mill. Total time diverted to the holding ponds was thirty-six hours fifteen minutes.

In October waste was diverted to the holding ponds thirty-seven hours, thirty minutes of which seventeen hours, thirty minutes were during a mill power failure and for repairs to the lime screw and replacement of the untreated waste flow nozzle with an eccentric orifice. The remaining twenty hours were due to high color.

With manual control of the lime feed system and the excessive diversion to the holding ponds, the operating results were better than expected. Lime concentration objective was 1200 ppm $Ca(OH)_2$. In September the average for the period was 1394 ppm and in October the average was 1266 ppm. Average lime consumption for the two periods in sequence as tons 90% CaO per million gallons of waste treated was 5.01 and 4.57. The average percent reduction in process waste color was 90.4 for September and 91.4 for October. Color reduction per 100 ppm $Ca(OH)_2$ for the range of untreated waste and lime concentration are shown in the Appendix, page 114 and page 115, for the two reporting periods.

Waste flows in million gallons were as shown in Table 11.

TABLE 11

	Uncontaminated Waste To Riceboro Creek	Treated Process Waste			
		From Mill	From Holding Pond	Total	Diverted To Holding Pond
Sept., 69	124.63	150.15	29.23	179.38	12.11
Oct., 69	112.90	129.17	25.19	154.36	8.33

Tests for COD were not run because of the heating shelf, required for this test, burned out. Material delivery and availability of repairmen delayed its replacement. Some of the BOD tests were

depleted of oxygen before completion. To correct this, a new glass lined water still was ordered. Test data on the completed BOD tests is shown below with corresponding color concentrations.

TABLE 12

BOD REDUCTION VS COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	UNTREATED BOD ppm	TREATED Color APHA Units	TREATED BOD ppm	% REDUCTION Color	% REDUCTION BOD
9/3	1000	236	130	225	87.0	4.7
9/4	1760	350	120	251	93.2	28.3
9/9	1760	421	170	281	90.3	33.2
9/9	1580	296	160	216	89.9	27.0
9/11	1020	289	160	213	84.3	26.3
9/18	2000	545	150	273	92.5	49.9
9/23	1000	253	80	246	92.0	2.7
Av.	1445	341	138	243	89.8	24.5
10/1	1860	486	140	361	92.5	25.7
10/7	2100	472	130	234	93.8	52.2
10/8	1360	313	125	208	90.7	33.5
10/14	1400	367	150	267	89.3	27.2
10/15	3080	607	150	264	95.1	56.5
10/16	1880	396	180	279	90.4	29.5
10/21	1060	312	110	293	89.6	6.1
10/22	1660	564	125	286	92.5	49.7
10/23	2000	693	170	361	91.5	47.9
10/28	1460	323	110	267	92.5	17.3
10/29	1540	462	100	279	93.5	39.6
10/30	1660	550	140	252	91.6	54.2
Av	1755	462	135	279	91.9	36.6

Operating Period - November 1 through December 31, 1969

Modifications to Equipment:

Specific gravity measurement of slaked lime from slaker.

The centrifugal pump and constant headbox were completed on schedule, November 5. This pump did not give the flow capacity required. As it did not come up to purchase agreement it was returned to the supplier for credit against the purchase of a Monyo pump. This was the third attempt to obtain a satisfactory

pump for this service.

Operating Results:

The severe corrosive and abrasive characteristics of lime have begun to show their effect upon the lime feed equipment after twenty months of service. The variable speed drive on the screw supplying the lime to the system has been a continuing source of trouble. It has been rebuilt twice and there are still problems in adjusting and maintaining an adequate range of screw speed.

Replacements include the slaker agitator motor, slaker reject rake bearings, slaked lime slurry agitator and motor, lime slurry pump impeller, lime slurry recirculation control valve, and two manual lime slurry dump valves. The lime slurry circulation pump for specific gravity measurement has been replaced twice and is to be replaced a third time. The lime slurry feed pump casing has been welded. The pump shaft is worn to such an extent that packing will not hold more than twnety-four hours. The pump shaft, impeller and casing, are scheduled for replacement.

In November repairs were made to the lime screw variable drive on the run,without diverting the waste to the holding pond,in an effort to hold treatment until the mill scheduled shutdown in December. On December 11 and 13 it was necessary to divert to the holding ponds for a total of twenty-two hours because of low lime feed and high untreated waste color. Because of this and the amount of work to be done, the lime feed system was shut down for maintenance and repairs on December 20. Not including the two-day mill shutdown, December 24 and December 25, untreated and treated waste were diverted a total of 104 hours and 20 minutes to the holding ponds in December.

Waste flows in million gallons were as follows:

TABLE 13

	Uncontaminated Waste to Riceboro Creek	Treated Process Waste From Mill	Hold. Pond	Total	Diverted to Holding Pond
Nov., 1969	122.05	172.95	30.30	203.25	00.00
Dec., 1969	112.02	143.24	8.79	170.39	29.39

Attempts to put the lime feed system on automatic control were

aborted by the malfunction of the variable speed lime screw drive. Adjustments to the fixed lime feed were on the basis of average process waste flow with an objective of 1200 ppm $Ca(OH)_2$ concentration to the floc tank. In November the average calcium hydroxide concentration was 1189 ppm. In December the average was 1272 ppm $Ca(OH)_2$. Average lime consumption for the two periods in sequence was 4.13 and 4.24 tons 90% CaO per million gallons of waste treated. The averag. feed rate of lime was close to the desired rate. However, the variations in feed were such to give the poorest performance in color removal since the plant has been in operation. The average percent color removal for November and December was 89.3 and 89.1. Color reduction per 100 ppm $Ca(OH)_2$ for the range of untreated waste and lime concentration are shown in the Appendix, page 116 and page 117 for the two reporting periods.

Tests for COD were not run in November because the heating shelf required for the test had not been repaired. BOD tests were made by the Interstate lab in November and by an outside commercial laboratory in December. Results are shown below. The shortage of lab work in December was due to the replacement and training of a lab technician and due to holidays in this period.

TABLE 14
BOD REDUCTION VS. COLOR REDUCTION
PROCESS WASTE

Date 1969	UNTREATED Color APHA Units	UNTREATED BOD ppm	TREATED Color APHA Units	TREATED BOD ppm	% REDUCTION Color	% REDUCTION BOD
11/4	2000	572	200	376	90.0	34.3
11/5	2000	573	200	351	90.0	38.7
11/6	1580	508	130	321	91.7	36.8
11/11	1760	371	130	305	92.6	17.8
11/12	1500	437	150	268	90.0	38.7
11/13	2000	725	175	427	91.3	41.1
11/18	1280	314	120	219	90.6	30.3
11/19	1100	257	120	208	89.1	19.1
11/20	2000	427	160	302	92.0	29.3
Av.	1691	465	154	309	90.8	31.8
12/4	1400	450	200	188	85.7	58.2
12/11	1600	408	120	160	92.5	60.7
12/18	2400	426	200	168	91.7	60.5
Av.	1800	428	173	172	90.0	59.8

Clarifier Solids Balance:

Calcium analyses of the clarifier feed, underflow and overflow, were made over the period of October 20, through November 21, 1969. The data is tabulated below.

TABLE 15

CLARIFIER CALCIUM BALANCE

		Calcium as $CaCO_3$			Calculated Values		
	Clarifier Feed Mil. Gals.	Clarifier Feed ppm	Clarifier Underflow ppm	Clarifier Overflow ppm	Clarifier Overflow Mil. Gals.	Clarifier Underflow Mil. Gals.	Ca in Over flow % of Feed
Date							
10/20	5.98	2600	13,240	1007	5.20	.78	33.7
10/31	7.47	1480	12,700	1093	7.22	.25	71.4
11/3	7.32	990	23,793	831	7.27	.05	83.4
11/5	7.44	2212	23,333	959	7.02	.42	40.9
11/6	6.80	1468	14,500	972	6.55	.25	63.8
11/7	7.12	1480	23,333	957	6.95	.17	63.1
11/10	5.65	2964	26,000	1008	5.07	.58	30.5
11/12	7.11	1976	17,167	1060	6.71	.40	50.6
11/13	6.62	1940	17,333	1104	6.28	.34	54.0
11/18	7.04	1612	46,600	972	6.94	.10	59.4
11/19	6.86	976	14,333	960	6.85	.01	98.2
11/20	6.52	1592	12,666	908	6.14	.38	53.7
11/21	6.40	1552	12,167	972	6.07	.33	59.4

Calcium was determined by titrating samples taken with 0.01 M EDTA (disodium salt of ethylenediamine tetraacetate) using HHSNN indicator. (Fisher Chemical Co., Cat. No. H-342). Potassium hydroxide (8M) solution was added to the samples, allowing five minutes for aging before titration. On November 3 and November 19, the calculated volumes of underflow were 3.5 and 17.5 percent, respectively, of the estimated flow obtained from the pump curves. These discrepancies can be attributed to the small sample volume of 0.3 ml used for titration. It was necessary to use this sample volume because of the high calcium concentration in the underflow.

The above data shows that the concentration of lime in the overflow is relatively constant regardless of the lime concentration in the feed. All lime added in excess of 1,000 ppm as $CaCO_3$ will leave the system as sludge in the underflow. At the level of lime treatment during the test work, total calcium in the overflow averaged 52.8%. If treatment level was maintained at 1200 ppm $Ca(OH)_2$, 61% of the calcium would be in the clarifier overflow.

C. Summary

Equipment Evaluation

Lift Pumps:

The lift pumps are the heart of the system. If they cannot keep up with the effluent flow, the mill has to shut down. Untreated process effluent cannot be diverted to Riceboro Creek under any circumstances.

The pumps initially installed were selected with price as a factor in an effort to cut cost. No savings were realized as these pumps had to be replaced by Rust Engineering Company within four months because of excessive wear and outage time for replacement of bearings, shafts, and impellers.

The Hazleton pumps now in service are for heavy duty at 875 rpm with forced water bearing lubrication. Impellers are 28% chrome alloy. Performance has been excellent,with no signs of wear after eighteen months service.

Chemical Feed System:

For a chemical waste treatment system to operate successfully under such varying conditions of flow as previously described, there must be a very flexible chemical feed system which will respond to accurate flow signals with a minimum of lag time. To accomplish this, the initial control of lime feed as a function of the output of the variable screw conveyor to the slaker was changed to the existing control of slaked lime slurry at a fixed concentration of calicum hydroxide. The eccentric orifice with pressure sensor oil seals replaced a flow tube.

The chemical feed system may not be the best that can be installed,but it does utilize all of the equipment that was available at startup. Attention has been given to details such as to provide constant water pressure at the control valves. A further improvement would be to maintain a constant temperature in the slurry tank. However, the system can and does maintain a relatively constant predetermined concentration of calcium hydroxide in the process waste to the floc tank.

Lime is an aggravating material to handle. Approximately

ninety percent of the maintenance cost for the entire waste treatment system is for lime handling.

Flash Mixing:

An inline flash mixer was provided. It was a continuing source of trouble,as a hard brown scale would build up on the impeller and internal baffles to a point where excessive vibration and head loss would occur. After modifying the internal baffles, conditions were somewhat improved.

To evaluate the need of flash mixing, the plant was operated for thirty-eight days with the inline agitator removed. Operating results were compared with selected periods having corresponding mill conditions. From this it was concluded that flash mixing does not affect the degree of color removal obtained. Baffling to break up laminar flow is adequate. Consequently, the inline agitator has been out of service since the first part of June, 1969.

Flocculation:

The floc tank was designed for fifty minute retention at a maximum waste flow of 10 mgd and has a double bladed turbine agitator with a variable speed drive of a range up to 14 rpm. Test work by The Rust Engineering Company indicated that the optimum flocculation time is between fifteen and twenty minutes. However, the longer retention time was provided for on the insistence of the Georgia State Water Quality Control Board in their approval of the construction drawings. Subsequent removal of the floc tank overflow funnel to reduce turbulence and the accumulation of solids has necessarily reduced the minimum retention time from fifty to thirty-five minutes.

A program to evaluate flocculation was aborted by the excessive buildup in the floc tank of lime mud and dregs from the mill recausticizing area with any agitation less than the maximum. By-passing the floc tank was not tried because of the real possibility of plugging the underground pipe line to the clarifier center feedwell. Because of the concern with this factor, the underground pipe line was opened and inspected after eighteen months service. There was no scale or buildup of solids. Apparently flocculation gives a side benefit of effluent conditioning.

Clarification:

The clarifier was sized on a rise rate of 0.24 gallons per minute per square foot of clear rise area (gpm/sq ft) for 10.0 mgd of effluent. At the present average flow of 5.48 mgd, the rise rate is 0.13 gpm/sq ft.

The clarifier is subject to considerable short circuiting because of the large exposed surface area at the elevated location. Absorbed calcium carbonate from the atmosphere causes calcium carbonate scale buildup in the launderer notchweirs and the discharge pipe line. However, solids carry-over is 10 ppm and less. Solids concentration in the underflow average 2.0 percent.

Because of the coagulation of lime and fiber, a very readily settleable conglomerate is formed. For this reason and for better flow characteristics, the clarifier could be conservatively designed for an average 0.50 gpm/sq ft rise rate.

Lime concentration in the feed to the clarifier affects the quality of the clarified waste and underflow. At treatment levels below 1,000 ppm calcium hydroxide, color begins to rise significantly with an increase in turbidity. With treatment levels above 1,000 ppm, the decolorized effluent from the clarifier has an average concentration of 722 ppm calcium hydroxide with a plus or minus deviation of 14 percent, irrespective of the lime treatment level. Consequently, there is an increase of free lime in the sludge underflow in direct relationship to the calcium hydroxide added. This can be as much as 1,200 percent with a 500 ppm change of lime concentration in the feed. Primary sludges from untreated kraft pulp mill effluent are difficult to dewater because of their hydroscopic characteristics. Addition of calcium hydroxide makes the problem of dewatering more difficult and adds problems to calcining in a calcium recovery process.

Natural Stabilization:

The decolorized and clarified effluent entering the natural stabilization lake is sterilized because of the high alkalinity at a pH of 12.2. The chemical properties are such that there are three reaction processes required to make it suitable for discharging to Riceboro Creek, namely, physical, chemical and biochemical.

In the first 100 acres of the lake, carbon dioxide absorbed from

the atmosphere precipitates practically all calcium as a carbonate and the pH drops to 10.2. Without the buffering effect of the calcium, the pH drops rapidly as the effluent moves on from this area. With the decrease in pH to 10.2, some color is released from previously deposited calcium. Color of the effluent increases as it flows on through the lake. This color is from natural origin and contributes to the color increase in the first 100 acre zone because of windage.

Biochemical oxygen demand of the effluent is reduced in two reaction processes. The first is by chemical oxidation with oxygen from the atmosphere which occurs in the first 100 acres of the lake and then by the biochemical reactions under aerobic conditions in the remaining part of the lake.

Mechanical aeration before discharge plays no part in BOD reduction. Its function is to assure a minimum of 6 ppm dissolved oxygen in the effluent.

The stabilization lake area could be reduced by as much as 140 acres to obtain the same quality of effluent with pre-calcium precipitation.

There are three points of particular note in the biochemical stabilization of kraft mill waste following lime treatment for color removal. Foaming is eliminated after calcium precipitation. This is a distinct advantage if an accelerated aeration process is to be used. Phosphorus is removed by precipitation, and nitrogen is converted to the ammonia ion which can be removed by stripping [8]. This is an advantage in the natural process as it makes control of eutrophication unnecessary. Sterilization at the high alkalinity makes it unnecessary to separate septic tank effluent.

Operating Results

Lime Treated Effluent Color:

Effluent color of lime treated kraft pulp mill effluents is affected to some degree by sodium alkalinity and by the color concentration in the untreated effluent.

Taking the background mill operating conditions into consideration, effluent treatment operating data presented is based on averages. Figure 3, page 59, shows the average treated waste color for a range of treatment of 1,000 to 2,400 ppm calcium hydroxide with average untreated effluent colors of 460, 840, 1210, 1610, and 2120 ppm. The averages are in grouping of 400 ppm of untreated color and represent 456 operating days of the first two years of operation. Data for days when there were mill upsets or when effluent was diverted to the holding lagoon are not included. Untreated effluent color is in APHA cabolt units at observed pH, which averaged 10.4. Treated effluent color is at an adjusted pH of 7.6.

Even with data averaging there is considerable scattering of points plotted in Figure 3, page 59. This possibly can be attributed to the uneven number of days for each point. However, there is a remarkable close conformity of slope of the lines drawn through these points for each level of untreated effluent color. The important relationships to be noted are that there is only approximately 2 ppm of color improvement in the treated effluent with each 100 ppm calcium hydroxide increase above 1,000 ppm; and that there is approximately 100 ppm increase with an increase of 1,700 ppm untreated effluent color.

Lime Treated Effluent COD and BOD:

Lime treatment also reduces chemical oxygen demand (COD) and biochemical oxygen demand (BOD) of kraft pulp mill process effluents. The overall average reduction through the clarifier at the Riceboro mill has been forty-six percent for COD and thirty-four percent for BOD_5 on unfiltered composite samples.

Figure 4, page 60, shows these values for the treated effluent compared to the treated effluent color. COD appears to be a straight line function whereas the BOD_5 is a well defined exponental function. In the lower ranges of treated color, BOD_5 values approach a minimum of 180 ppm.

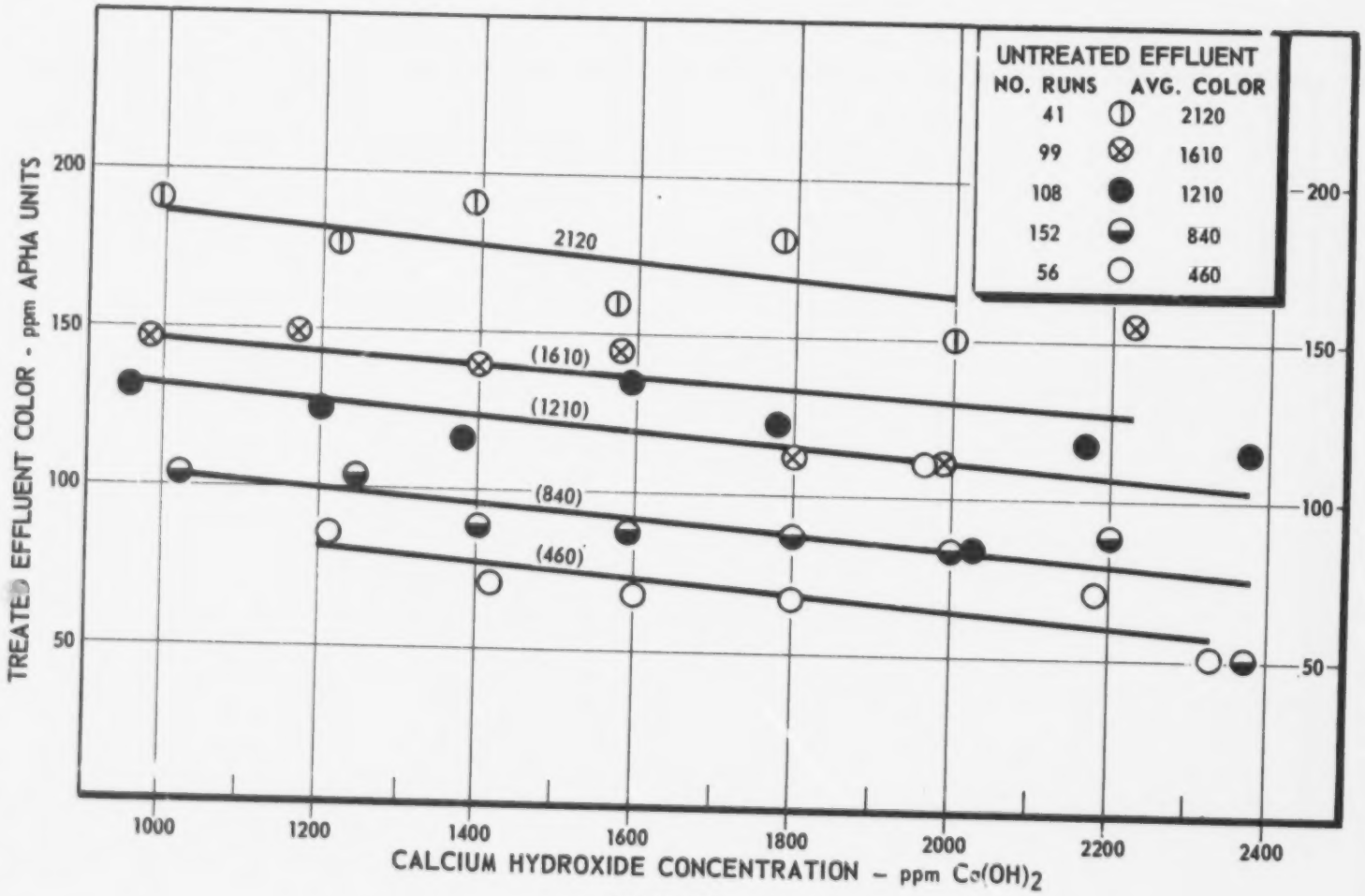


FIGURE 3
TREATED EFFLUENT COLOR vs LIME CONCENTRATION

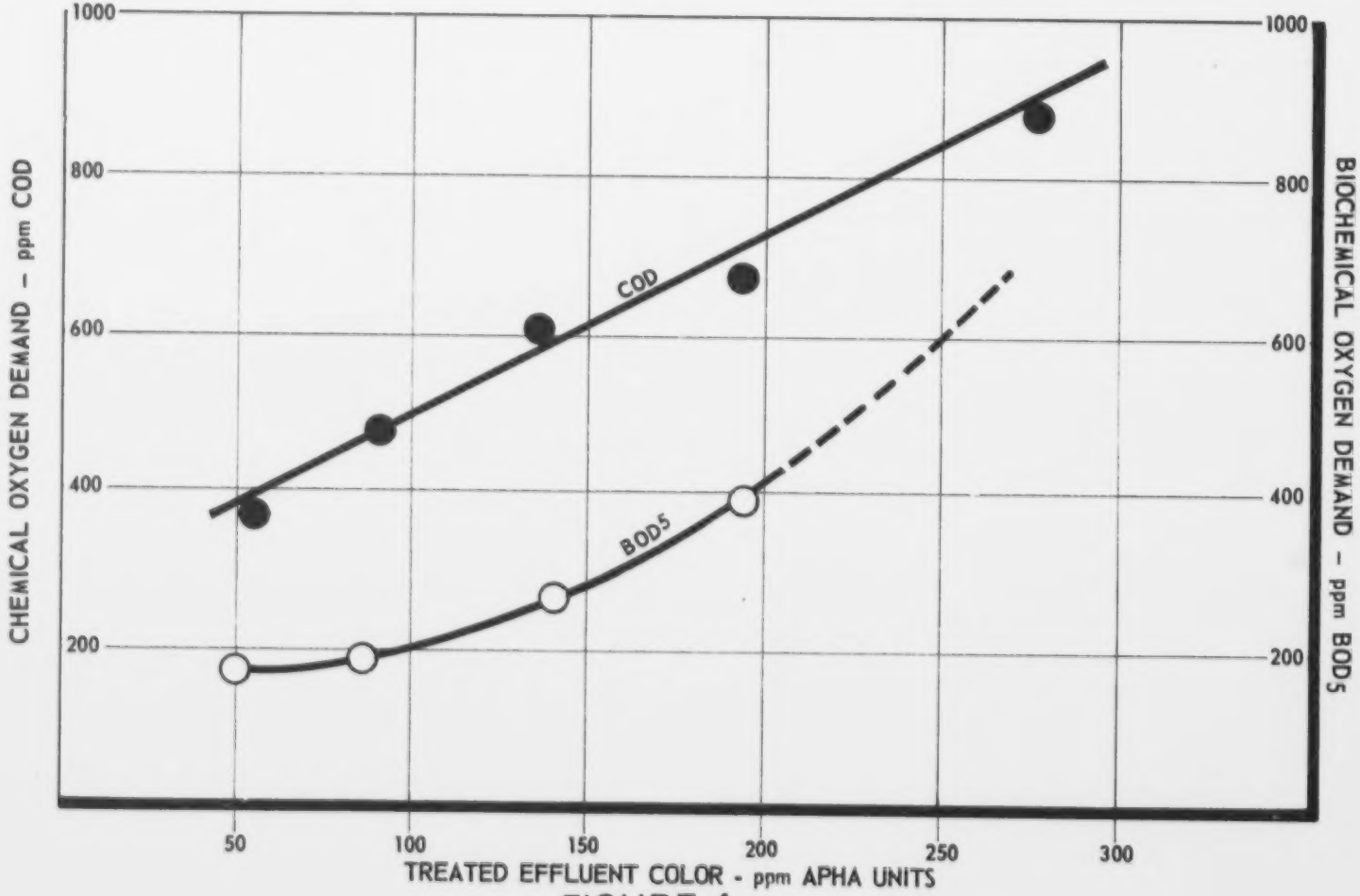


FIGURE 4

TREATED EFFLUENT COLOR cs. vs COD and BOD_5

Natural Stabilization of Lime Treated Effluent

Figure 3 page 62 shows the color and BOD_5 profile of the lake in relation to the number of acres in line of flow. The data obtained from weekly surveys, covers a period of one year from March of 1969 through March of 1970. During this period the overall lake color has continued to decline.

The zone of natural calcium precipitation is indicated by the shaded area to the left of the chart. In this area the color increases approximately 40 ppm. In the biochemical stabilization area, color increase rapidly and then as a straight line function in relation to the number of acres covered. It is presently not known whether or not there is a further reduction of the process effluent color in the final stages of stabilization. Research work at the FWQA, Southeastern Laboratory in Athens, Georgia, may give the answer.

The decrease of biochemical oxygen demand following calcium precipitation follows the trend of maximum performance for a natural biodegradation system.

In summary, overall chemical treatment in terms of process color is 90% and biochemical treatment is 98%. Following is a comparison of average mill discharge with regulatory limitations as imposed expressed in lbs per day. The limitation on color in lbs per day is based on 30 ppm color and a maximum flow of 20 mgd as defined in the mill construction permit.

TABLE 16

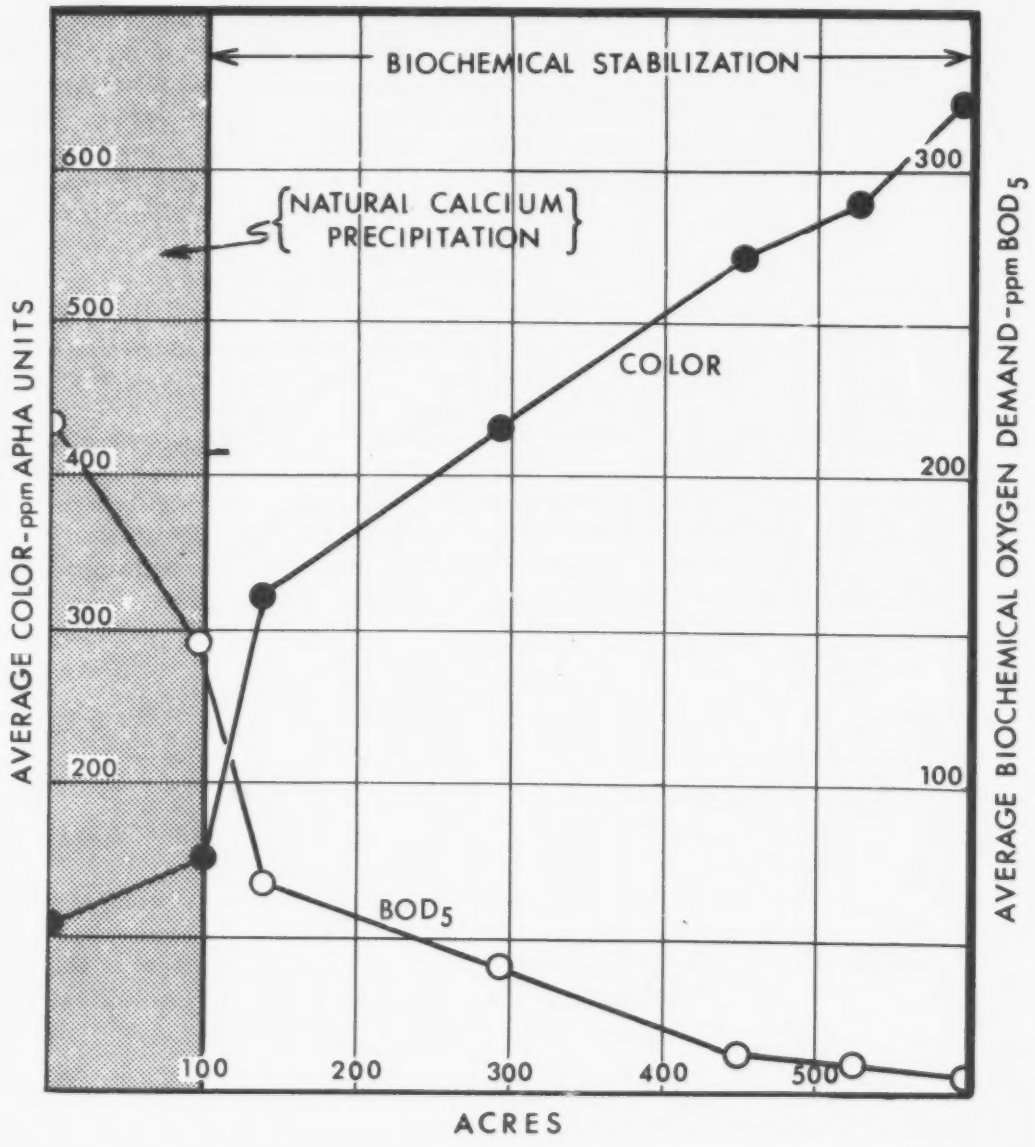
	Actual #/D	Limitation #/D
Color	5,020	2,500
BOD	485	800

Because of the color contributed from natural sources within the biochemical stabilization lake, waste discharges were restricted to a five percent dilution with fresh water inflow at Half Moon Landing, based on charts furnished by U. S. Geological Survey office in Atlanta.

The quality of the treated effluent discharged is comparable to Riceboro Creek waters with the exception of dissolved inorganic

FIGURE 5
STABILIZATION LAKE COLOR & BOD PROFILE

March 1969 - March 1970



solids. Color at times is lower than the creek water. BOD averages 6 ppm with dissolved oxygen at saturation. Because of the effluent stabilization there is no oxygen sag in Riceboro Creek below the mill.

Lime Treatment Cost

The actual cost of equipment and facilities for lime treatment came to $355,100 exclusive of the clarifier and holding lagoons. This figure can only be used as a guide because of the possible reduction in equipment requirements outlined in the commentary on facilities and because of the escalation of equipment cost.

Chemical lime is the major operating cost. Figure 6, page 64 shows the data of Figure 3 in terms of percent color reduction compared to the cost of chemical lime per million gallons of effluent treated.

Lime is charged to the treatment plant at $15.35 per ton as ninety percent calcium oxide. With untreated effluent colors above 460 average, the increase in percent reduction is approximately 2.0 with an increase of 1,400 ppm calcium hydroxide. This gives a cost of $38 per million gallons for a one percent change in color above the 1,000 ppm calcium hydroxide treatment level.

Labor listed below is in terms of hours per year for conversion to current cost at any locality.

Operating Labor	8,760 Hr/Yr
Maintenance Labor & Supervision	6,500 Hr/Yr

The above figures are for the entire waste treatment system. Approximately eighty-five percent of the labor is for lime treatment.

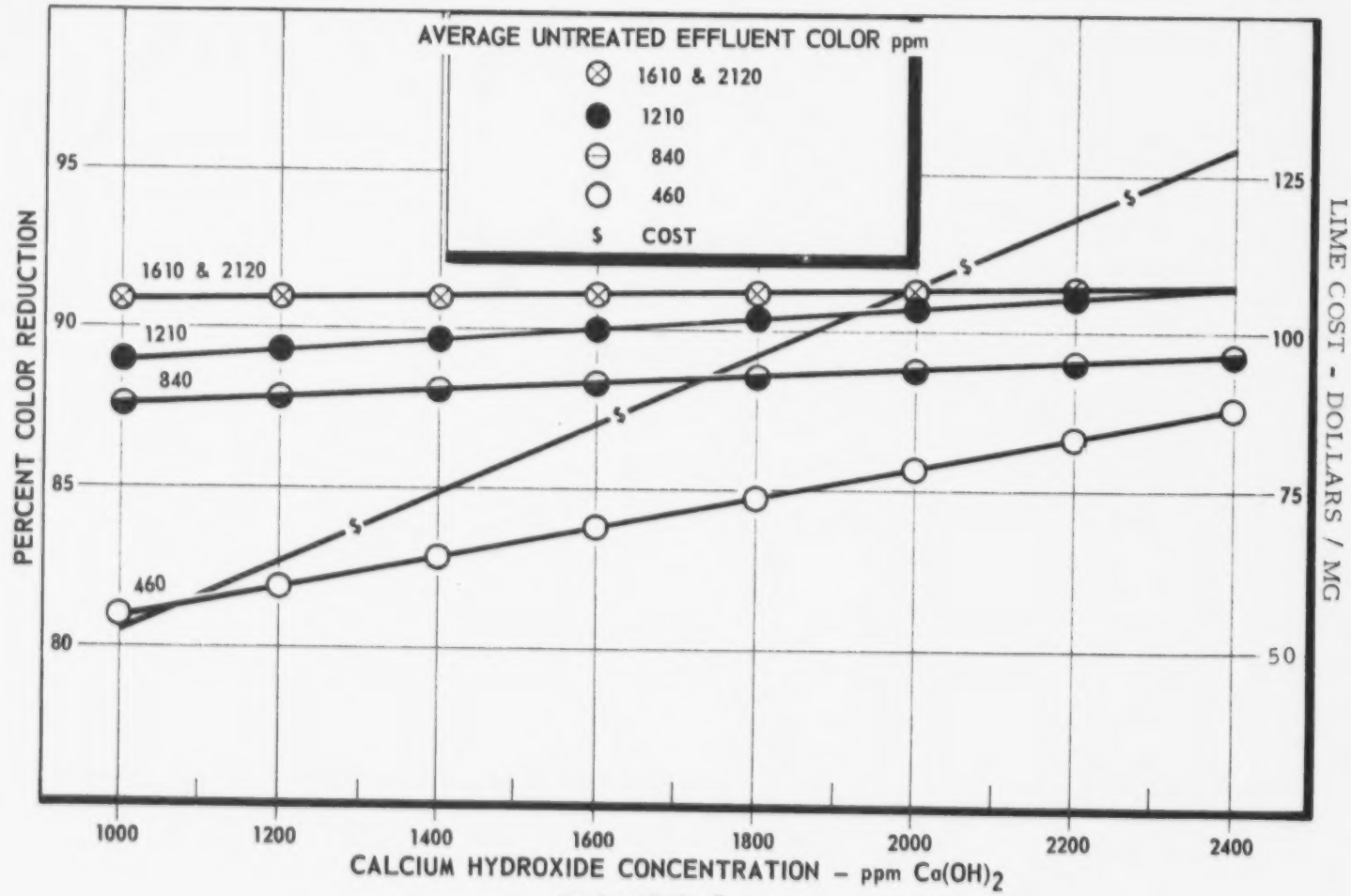


FIGURE 6

PERCENT COLOR REDUCTION AND COST vs LIME CONCENTRATION

SECTION VI

CALCIUM RECOVERY - MATERIALS AND METHODS

A. Carbonation Process

The carbonation process can best be described as a physical process combined with a chemical reaction process.

The transfer of carbon dioxide from a gaseous to a solution phase is the first step in the physical process. The rate of transfer is a function of bubble size, time of contact, temperature and gas solubility. As the carbon dioxide goes into solution, it reacts with the dissolved calcium hydroxide to form calcium carbonate. Within pH ranges of 8.0 and 10.5, calcium is almost entirely in the insoluble carbonate form [9].

The succeeding physical process is the crystalline growth of the precipitated calcium carbonate particles. In a supersaturated solution, precipitation can occur in two ways - by spontaneous self nucleation, or by the deposition on solid material. In the first instance, a large number of colloidal particles are formed which do not settle. This phenomenon occurs at a high degree of saturation such as exists with the lime treated kraft waste. In general, the rate of precipitation is proportional to both the degree of supersaturation of the solution and the available surface of previously formed particles. Time is required for particle growth to take place [10].

The final phase of the physical process is the particle settling and thickening into a slurry or sludge. Calcium carbonate does not follow the reduced solids concentration flux curve for rigid spheres. Within well defined concentrations, calcium carbonate goes through a period of free settling followed by aggregation of particles and finally, compaction settling to maximum density [11].

There is an overlapping of both physical and chemical processes except in the final phase of compaction settling. To design equipment for the process, the system was divided into four sections: carbonation, flocculation, settling, and sludge concentration.

B. Pilot Plant

Design

The continuous pilot plant at Interstate Paper Corporation was designed to be able to control and vary all factors which would affect the process. A carbonator was provided for the formation of calcium carbonate particles by the physico-chemical and chemical reaction process. A separate unit was installed for the physico-chemical process of flocculation settling, and concentration.

The carbonator was built for this specific application. The second unit is a standard Densator pilot plant obtained from the Fuller Company/General American Transportation Corporation on a rental basis. Cross sectional sketchs of both units are shown on Figure 7, page 67, and Figure 8, page 67. Overall dimensions are: Carbonator 30 inches ID by 15 feet; Densator 42 inches ID by 10.5 feet.

Instrumentation was selected so as to have the maximum application in a future full scale plant and to enable the evaluation of their capabilities for control. Controls are the electronic type and were obtained from Foxboro.

Auxiliary equipment, including pipe, valves, and pump was sized to provide a plant hydraulic capacity of 43,200 gallons per day. Carbon dioxide gas supply was from two one ton dry ice converters. Air was from mill supply.

Figure 9, page 68 , shows the flow diagram of the pilot plant. Included are the modifications to the gas supply system, carbonator piping system for counter flow of water and gas, and slurry withdrawal from the Densator.

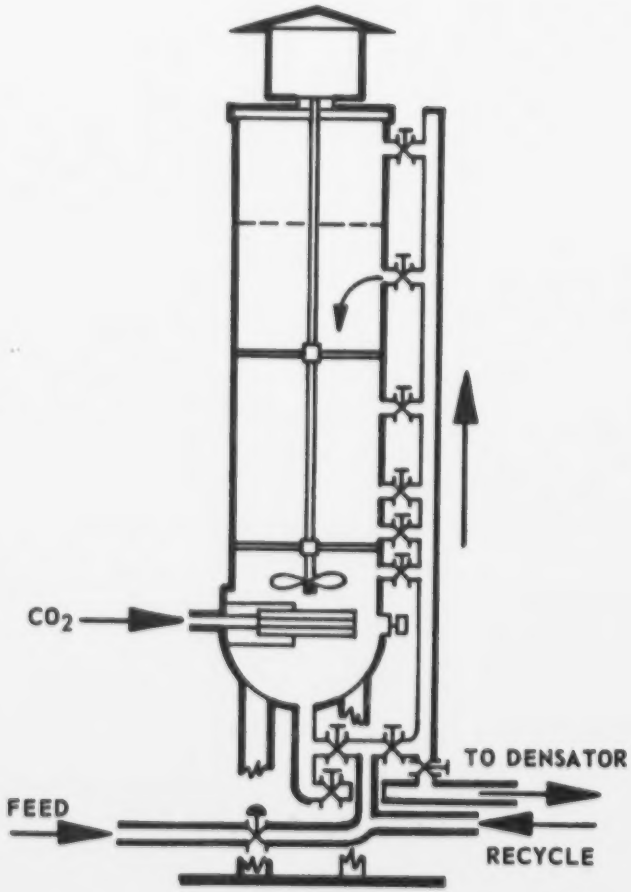


FIGURE 7
CARBONATOR

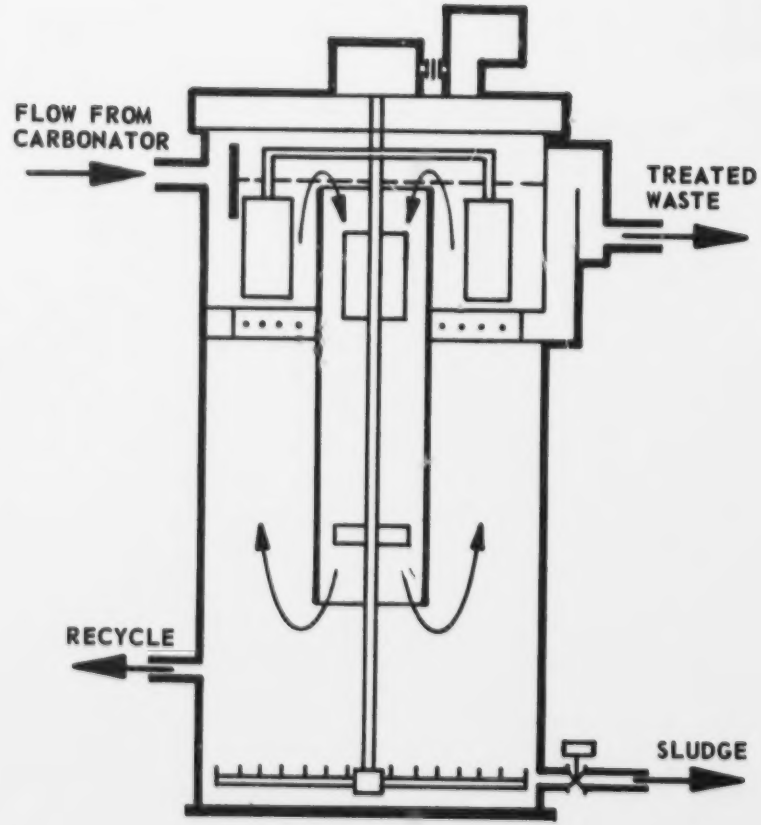


FIGURE 8
DENSATOR

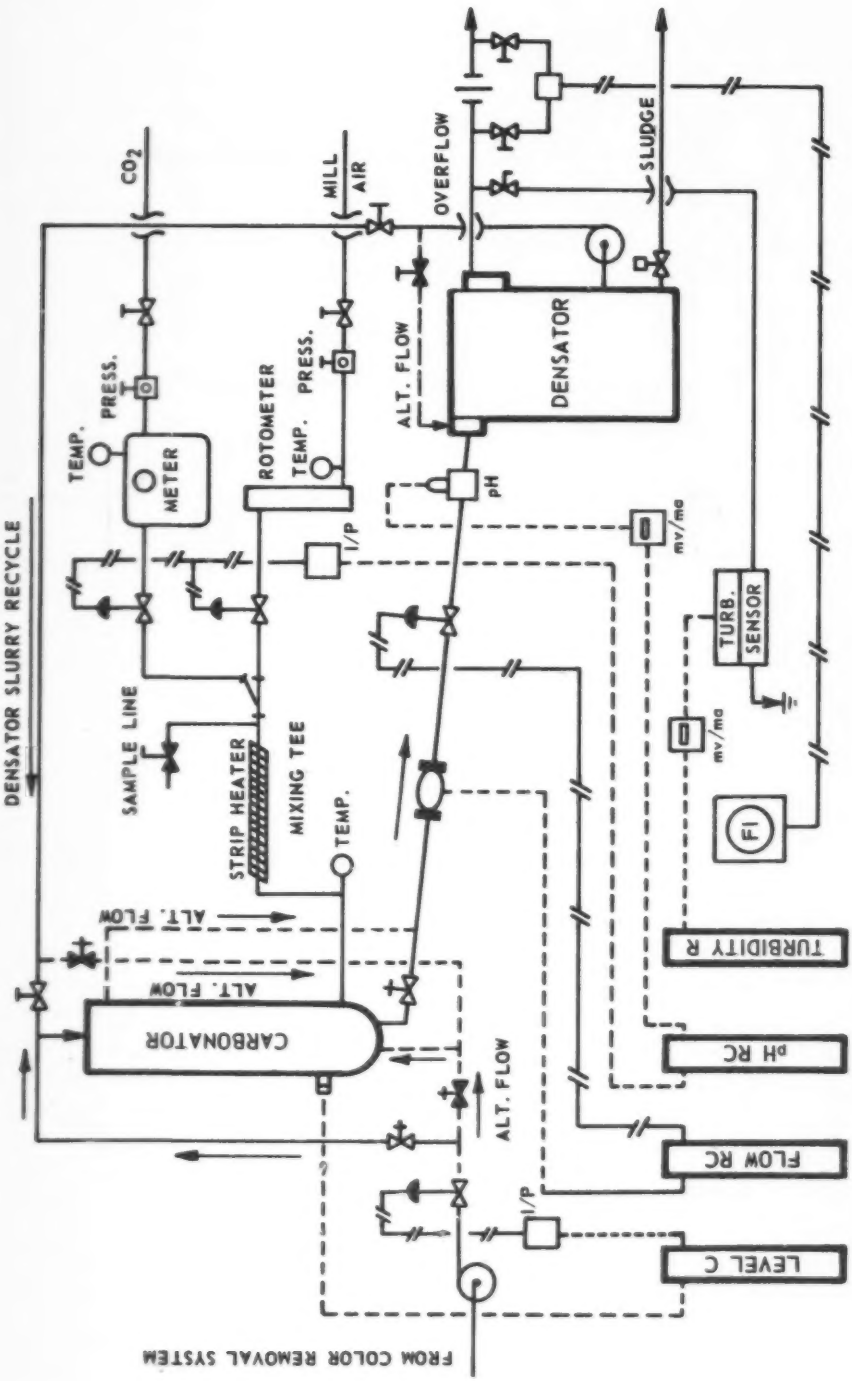
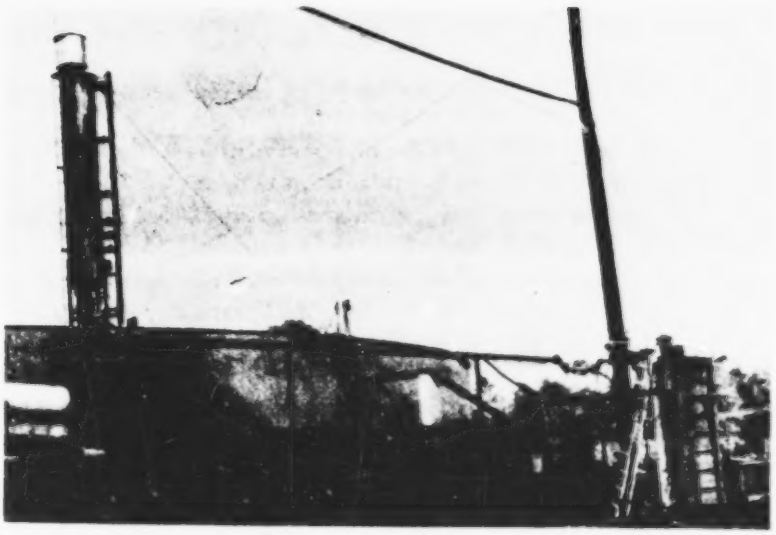


FIGURE 9
PILOT PLANT FLOW DIAGRAM

Layout and Orientation

Figure 10



In the above photograph, the mill's 200 ft. diameter primary clarifier is shown in the background, with the sloped 36 inch overflow line in the foreground. The tall elevated tank with the perpendicular manifold to the left is the carbonator. The two horizontal cyclinders in the left foreground are the dry ice storage for the carbon dioxide system. To the far right is the Densator. The small, sloped, elevated pipe line connects the carbonator and Densator. In this line just ahead of the Densator, is the magnetic flow sensor, flow control valve and pH electrodes. The horizontal, small pipe line below the magnetic flow sensor is the overflow from the Densator. Just below this line are the flow-through reference cell, turbidity meter, and pH indicator mounted in the enclosed panel board. Control and recording instruments, motor switch gear, and field laboratory are in a small building to the right of the Densator. This building is not shown in the above picture.

In the order of water flow, lime treated mill process effluent is withdrawn from the primary clarifier overflow line by a pump mounted on the base below the carbonator. Flow is then into the perpendicular manifold to the carbonator at one of the selected levels and then out of the base of the carbonator by gravity flow through the sloped, small pipe line to the Densator. Flow is

controlled by the automatic valve ahead of the Densator. A d/p cell mounted on the lower part of the carbonator shell and a control valve on the discharge side of the pump are part of the system that maintains a set water level in the carbonator.

A pump at the base of the Densator returns partially settled calcium carbonator slurry through a rubber hose back to the lime treated effluent feed line to the carbonator. An in-line booster pump mounted on the carbonator elevated platform increases flow as needed. Heated carbon dioxide and air enter the carbonator near the bottom. Not shown is a rectangular removable plate to facilitate inspection and interchanging of diffusers. The cylindrical housing on top of the carbonator is a cover for the variable speed motor on the carbonator agitator.

The settled carbonate slurry is discharged by gravity to collection drums behind the Densator. Control is by a timer operated solenoid valve.

The flow pattern outlined above is the one used for the experimental runs. Referring back to Figure 9 page 68, it will be noted that there are several variations. Water flow can be concurrent to gas flow and withdrawal from the manifold. Settled slurry can be recirculated back to the flocculation zone of the Densator, or a split flow of slurry can be used.

Startup and Shakedown

Each programmed test run was scheduled to have a stabilization period of not less than fifteen hours with all systems functioning properly. Change-overs were to be made by 5:00 P. M. each operating day with the test work starting at 8:00 A. M. the following morning for a period of one hour. Equipment maintenance and modifications were to be done by the mill maintenance department on the first shift.

On starting up, the first problem was with the gas control system. As originally planned, air and CO_2 were to be blended in a mixing tee, heated and controlled by one valve actuated by the pH controller. Differences in gas and air supply pressure gave a wide swing in carbon dioxide concentrations. This was resolved by installing a separate control valve for the air supply. Both valves were actuated by the pH controller. The mixing tee was relocated down stream from the control valves with the heating

tapes next in line. To increase gas transfer, piping changes were made for counter flow of gas and waste.

There were also difficulties with the carbon dioxide gas pressure regulators freezing, particularly so in extreme cold weather. Additional heating tapes finally gave fair control. However, a thermostat would have been of great help.

Sludge recycle to the carbonator was difficult to control. Both the pump suction and blow-down solenoid valve were on the same manifold connected to a twelve inch riser from the bottom of the Densator. The riser would plug as the density of the slurry increased to the desired consistency. First, the booster pump was added to increase and maintain the slurry flow. Then, it was found that the settled slurry was being transferred to the carbonator where it was building up and not getting out of the system through the Densator blow-down.

With the slurry recycle problem, there was also a severe foaming problem in the carbonator. At first it was thought to be caused by abnormally high organic color following mill upset conditions. Other possible explanations were a high solids to air ratio, low gas temperature and high pH. To eliminate the first possibility, the slurry recycle pump suction was changed to a point 18 inches above the bottom of the Densator. Next, the slurry blow-down riser and manifold were eliminated as plugging at this point became more severe.

Practically all of the problems with slurry recycle and foaming were eliminated with the above changes. Counter flow of waste and gas in the carbonator, higher gas temperatures and operating at the lower pH range of 9.0 apparently improved performance. Because of the difficulties encountered, many test runs were not acceptable. It was not until November 4, 1969 that the test program actually got under way. This was thirty days behind schedule. The delay was not a complete loss as much was learned about the process from observations.

Test Procedure

Following the plant stabilization period, all charts were checked for consistency of control. If satisfactory, three sets of samples of feed and Densator overflow were collected at 20 minute intervals. Accumulated slurry in blow-down barrels was measured and sampled. A composite sample of the feed was also taken.

Analytical and test procedures were as follows:

1. Calcium

   a. Soluble calcium concentration of the samples were determined by titrating with 0.01 M EDTA (ethylene diamine tetracetate) after the addition of 8 M potassium hydroxide and aging for five minutes. HHSNN (2-Hydroxy- 1-(2-Hydroxy-4Sulf-1-Naphthyl -AZO-3-Naphthoic Acid) was used as an indicator.

   b. Total calcium was obtained by following the above procedure after all calcium had been solubilized by acidification to a pH of about 4.0.

   c. Suspended calcium was by difference of total and soluble calcium.

2. Carbon Dioxide for Sodium Alkalinity

   Composite samples of feed were titrated with N/50 H_2SO_4 to the pH end point of the carbonated effluent using a pH meter. This gave the total alkalinity reacted express as calcium carbonate. Calcium alkalinity reacted was obtained by the decreased calcium in solution, also expressed as calcium carbonate. The difference between these two figures was converted to CC_2 and expressed as a percent of the total CO_2 used for sodium alkalinity.

3. Percent Carbon Dioxide in Gas Feed

   Percent carbon dioxide was calculated from the measured CO_2 and air volumes, temperatures and pressures. A Fyrite CO_2 indicator was used for verification.

4. Densator Slurry Underflow

   a. Specific gravity was obtained by weighing a 1,000 ml sample.

   b. Calcium content was obtained by titration using the procedure for calcium above.

5. Filter Leaf Test

An Eimco Filter Leaf Kit [18] was used to make filtration tests on the Densator slurry. The 1/10 square foot filter was covered with an Eimco polyethelene Style No. P O - 801 RF monofilament wire, thread count 105 x 40. Most tests were made with a formation time of 15 seconds and a drainage time of 15 seconds. In some instances, the formation time had to be reduced. All tests were made at 15 inches vacuum, unless otherwise noted.

There was one deviation from the Eimco procedure - there was no drying time. Thus, a test made with the above conditions was calculated to have 120 cycles per hour. Each filter rate then was multiplied by a factor of 0.8 as recommended by Eimco.

6. Slurry Carbonate Purity

Slurry carbonate purity was determined by analysis in the Continental Can Company's research laboratory. Calcium was determined by precipitation as the oxalate and carbon dioxide was measured by loss of weight on ignition.

C. Methodology Used in Computer Program

Program

A statistically designed experminental program based on the function block diagram, Figure 11, page 74, was used to minimize the number of experimental runs and yet gain maximum design and operation information.

A two-level factorial design of sixteen (16) runs was used to examine four design and operating parameters. Table 17, page 75, shows this experimental program and the performance response measured and studied. These parameters and responses were:

Parameters

Code		
1)	pH	Carbonation pH, dimensionless.
2)	Depth	Carbonation depth, in feet.
3)	Rate	Carbonation feed rate, in gallons per minute.
4)	Cycle	Densator recycle rate, in gallons per minute.

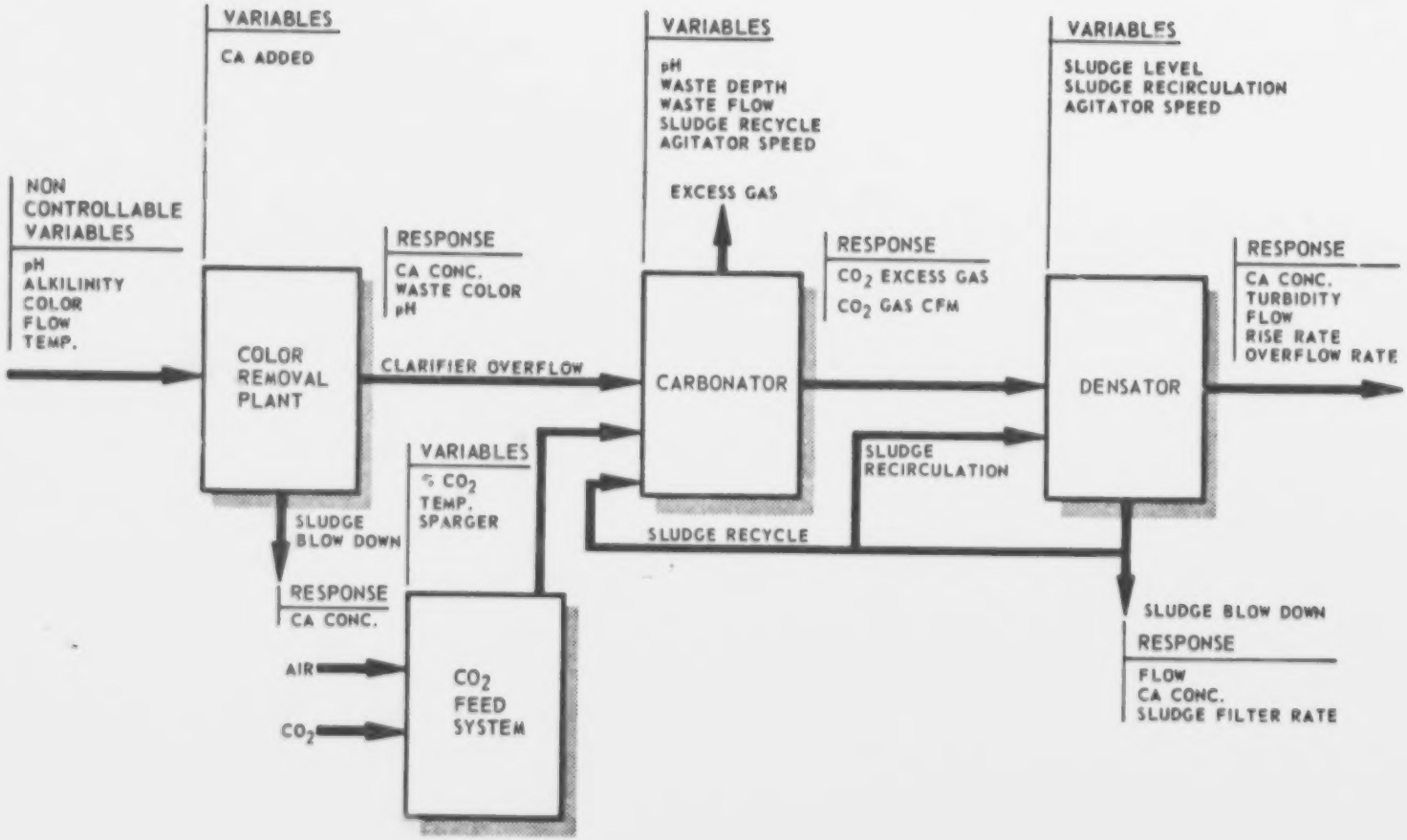


FIGURE 11

FUNCTION BLOCK DIAGRAM CARBONATION PILOT PLANT

TABLE 17

CARBONATION EXPERIMENTAL PROGRAM

Run #	PARAMETERS pH	Depth	Feed Rate	Recycle Rate	Responses To Be Measured
	+ = 10	+ = 10	+ = 20	+ = 3	
	- = 9	- = 5	- = 10	- = 0	
1.	-	-	-	-	CO_2 Transferred to $CaCO_3$
2.	+	-	-	-	Total CO_2 Transferred
3.	-	+	-	-	Dissolved Calcium loss
4.	+	+	-	-	Suspended Calcium loss
5.	-	-	+	-	Total Calcium loss
6.	+	-	+	-	
7.	-	+	+	-	
8.	+	+	+	-	
9.	-	-	-	+	
10.	+	-	-	+	
11.	-	+	-	+	
12.	+	+	-	+	
13.	-	-	+	+	
14.	+	-	+	+	
15.	-	+	+	+	
16.	+	+	+	+	

Notes:

1. Run experimental program in radom order.

Responses

	Code	
1)	CO-CA	CO_2 transferred to $CaCO_3$, in percent.
2)	CO-TO	Total CO_2 transferred, in percent.
3)	CA-DI	Dissolved calcium loss, in percent.
4)	CA-SU	Suspended calcium loss, in percent.
5)	CA-TO	Total calcium loss, in percent.
	CO-XX	2) - 1)

Table 18, page 77, shows a list of other feed and operating variables which were monitored during the experimental program and included in the computer study.

Tables 19, 20, 21, and 22, pages 78, 79, 80, and 81, show the experimental data generated by the sixteen (16) run statistically designed experimental program and duplicate runs during check-out of the equipment.

Data Analysis System

Certain key assumptions regarding the normality of the above raw data were checked to assure a proper interpretation of the calculated statistics. Although the number of degrees of freedom (total observations) available for calculating statistics and their confidence limits was adequate, the design used guaranteed orthogonality (independence) of the parameters.

The data collected from the above statistically designed experimental program and duplicate runs were analyzed, using a computer-based Data Analysis System [15]. This system is a coordinated set of user-oriented computer programs to statistically extract information from data, to develop mathematical models and to test their validity. It has available the usual mathematical modeling routines of simple correlation, simple and multiple regression. Also provided is an effective tear-down regression [16] and the new powerful heuristic regression [17, 18].

Study Results

All of the experimental data were studies in total. That is, twenty-one runs, four parameters, twenty-one other variables

TABLE 18

LIST OF OTHER VARIABLES MEASURED

FEED STREAM CHARACTERISTICS

	Code	
1.	FD-PH	Feed stream pH, dimensionless.
2.	FD-TP	Feed stream temperature, in degrees Fahrenheit.
3.	COLOR	Feed stream color, in APHA units at 7.6 pH.
4.	FD-DI	Feed stream dissolved calcium as $CaCO_3$, in pounds per 1000 gallons.
5.	FD-SU	Feed stream suspended calcium as $CaCO_3$, in pounds per 1000 gallons.
6.	FD-TO	Feed Stream total calcium as $CaCC_3$, in pounds per 1000 gallons.

CARBONATOR OPERATING VARIABLES

7.	ST-IR	Carbonator agitator speed, in revolutions per minute.
8.	Flow	Densator recycle flow, as percent of feed.
9.	SOLID	Densator recycle solids, as percent of feed.
10.	CO_2	Gas feed CO_2 concentration, in percent.
11.	TEMPC	Gas feed temperature in degrees Fahrenheit.
12.	TIME	Carbonator retention time, in minutes.

DENSATOR OPERATING VARIABLES

13.	*	Densator agiator speed, in revolutions per minute.
14.	FLOC	Densator floculation time, in minutes.
15.	RISE	Densator rise rate, in gallons per square foot per minute.
16.	TEMPD	Densator temperature, in degrees Fahrenheit.
17.	DE-DI	Densator overflow dissolved calcium as $CaCO_3$, in pounds per 1000 gallons.
18.	DE-SU	Densator overflow suspended calcium as $CaCO_3$, in pounds per 1000 gallons.
19.	DE-TO	Densator overflow total calcium as $CaCC_3$, in pounds per 1000 gallons.
20.	SP-GR	Densator underflow specific gravity.

SYSTEM TIME EFFECT

21.	DAYS	Time, as days since start-up.

* Held constant.

TABLE 19
CARBONATION PILOT PLANT OPERATING CONDITIONS

	Feed						Carbonator								Densator				
				Ca as $CaCO_3$ lbs / 1000 gal.				Densator Recycle		Gas Feed						Overflow Ca as $CaCO_3$ lbs. /1000 gal.			
Run No.	pH	Temp. °F.	Color - APHA @ 9.6 pH	Dissolved	Suspended	Total	Agitator - rpm	Flow % of Feed	Solids % of Feed	% CO_2	Temp. °F.	Retention - Min.	Agitator - ppm	Floculation -min	Rise Rate Gal/sq. ft./min.	Temp. °F	Dissolved	Suspended	Total	Under Flow Specific Gravity
1	12.2	95	180	8.677	.067	8.746	64	0	0	21.0	50	20.3	10	11.7	1.28	-	.634	1.068	1.702	1.1650
2	12.1	103	140	7.844	.237	8.078	61	0	0	28.5	162	23.4	10	12.5	1.19	100	.501	.926	1.427	1.1246
3	12.2	70	180	8.487	.184	8.670	64	0	0	14.3	350+	38.8	10	11.7	1.28	62	.492	.117	.609	1.1010
4	12.2	70	200	8.045	.100	8.145	63	0	0	24.2	187	38.8	10	11.7	1.28	62	.467	2.036	2.504	1.1649
5	12.0	98	140	10.148	.067	10.214	64	0	0	19.8	290	11.1	10	6.4	2.34	94	.467	.175	.643	1.1142
6	12.2	104	200	8.211	.526	8.737	64	0	0	50.5	97	10.7	10	6.1	2.43	100	.526	1.502	2.028	1.1146
7	12.2	94	140	9.238	.042	9.288	64	0	0	16.2	298	21.3	10	6.4	2.33	92	.459	.334	.793	1.0190
8	12.2	63	130	7.677	.559	8.153	64	0	0	36.5	119	20.3	10	5.8	2.56	63	.534	1.227	1.761	1.1280
9	12.1	102	175	9.113	.100	9.213	63	12.2	-	15.8	-	18.4	10	10.6	1.26	97	.100	.601	.701	1.0689
10	12.2	98	200	8.045	.067	8.111	64	20.0	-	18.8	61	16.9	10	9.7	1.28	96	.300	3.104	3.405	1.0579
11	12.1	99	180	8.011	.200	8.211	64	24.3	-	21.5	80	23.7	10	10.1	1.19	96	.300	1.769	2.070	1.0704
12	12.1	97	160	7.477	.100	7.577	63	45.0	-	17.2	169	24.2	10	8.0	1.28	86	.167	1.035	1.202	1.0659
13	11.9	102	120	8.045	.067	8.111	63	10.8	131	27.6	163	9.4	10	5.4	2.49	99	.234	1.302	1.535	1.0769
14	12.2	106	300	8.645	.234	8.879	64	14.3	-	18.3	350+	8.5	10	4.9	2.69	104	.476	2.328	2.804	1.0110
15	12.1	102	180	7.611	.067	7.677	64	23.1	-	14.7	-	16.2	10	4.9	2.50	100	.401	1.836	2.236	1.1175
16	12.2	100	120	8.946	.134	9.079	64	20.0	-	17.6	-	16.2	10	4.9	2.56	97	.434	.868	1.302	1.0902

TABLE 20
RESULTS OF STATISTICALLY DESIGNED
CARBONATION EXPERIMENT

Experiment		Variables				Responses				
						CO_2 Transfer %		Calcim Loss %		
Run No.	Date	pH	Carbonation Depth - Ft.	Feed - gpm	Recycle - gpm	To $CaCO_3$	Total	Dissolved	Suspended	Total
1.	12/12/69	8.7	5.0	10.00	0	20.5	22.0	6.9	11.6	18.5
2.	11/26/69	10.2	5.4	9.30	0	19.9	24.0	5.6	10.4	16.0
3.	12/23/69	8.7	10.0	10.00	0	27.1	34.7	5.6	1.3	6.9
4.	11/21/69	10.0	10.0	10.00	0	16.3	20.8	5.7	24.9	30.6
5.	1/7/70	9.0	5.0	18.25	0	21.0	26.4	4.2	1.6	5.8
6.	11/25/69	10.3	5.0	19.00	0	20.6	24.6	6.0	17.1	23.1
7.	1/6/70	8.7	10.0	18.50	0	25.6	33.0	4.9	3.6	8.5
8.	11/24/69	10.4	10.5	20.00	0	25.9	33.2	6.5	15.0	21.5
9.	11/13/69	8.8	5.0	9.80	1.20	20.0	25.5	1.1	6.5	7.6
10.	12/10/69	10.2	5.0	10.00	2.00	31.7	33.7	3.7	38.0	41.7
11.	12/9/69	8.8	10.0	9.25	2.25	35.5	46.0	3.6	21.3	24.9
12.	11/20/69	10.2	9.0	10.00	4.50	38.7	47.8	2.2	13.6	15.8
13.	11/18/69	8.9	5.0	19.40	2.10	20.4	25.5	2.9	16.0	18.9
14.	12/19/69	10.2	5.0	21.00	3.00	26.1	27.8	5.3	26.1	31.4
15.	12/8/69	8.8	10.0	19.50	4.50	33.4	42.1	5.2	23.7	28.9
16.	12/11/69	10.0	10.0	20.00	4.00	42.3	45.0	4.8	9.5	14.3

TABLE 21
CARBONATION PILOT PLANT OPERATING CONDITIONS
Duplicate Runs

Feed							Carbonator						Densator							
				Ca as $CaCO_3$ lb. s/1000 gal.				Densator Recycle		Gas Feed							Overflow Ca as $CaCO_3$ lbs. /1000 gal.			
Run No.	pH	Temp. °F.	Color -APHA @ 9.6 pH	Dissolved	Suspended	Total	Agitator -rpm	Flow % of Feed	Solids % of Feed	% CO_2	Temp. °F.	Retention-Min.	Agitator - ppm	Floculation-min.	Rise Rate Gal/sq ft./min	Temp °F	Dissolved	Suspended	Total	Underflow Specific Gravity
9	12.1	102	175	9.113	.100	9.213	63	12.2	263	15.8	-	18.4	10	11.6	1.26	97	.100	.601	.701	1.0689
9A	12.0	100	150	8.779	.067	8.846	63	10.0	123	15.2	-	18.4	10	10.6	1.28	94	.134	.534	.668	1.0260
9B	12.1	99	240	8.378	.033	8.412	64	26.5	141	23.2	145	19.3	10	14.1	1.06	97	.050	.618	.668	1.0522
9C	12.1	101	160	7.936	.050	7.986	64	21.1	110	-	191	17.6	10	12.3	1.22	93	.200	.993	1.193	None
9D	12.3	97	130	8.111	.100	8.211	33	25.6	3000	15.5	185	15.0	10	10.9	1.38	90	.159	1.344	1.502	1.0526
9E	12.2	98	200	7.953	.050	8.003	35	38.9	3600	22.2	180	16.2	10	13.0	1.15	86	.058	1.911	1.969	1.0658

TABLE 22

RESULTS OF STATISTICALLY DESIGNED
CARBONATION EXPERIMENT
Duplicate Runs

Experiment		Variables				Responses				
						CO_2 Transfer %		Calcium Loss %		
Run No.	Date	pH	Carbonation Depth-Ft.	Feed-gpm	Recycle-gpm	To $CaCO_3$	Total	Dissolved	Suspended	Total
9	11/13/69	8.8	5.0	9.80	1.20	20.0	25.5	1.1	6.5	7.6
9A		8.8	5.0	10.00	1.00	18.7	24.1	1.5	6.1	7.6
9B	11/10/69	8.7	5.0	8.30	2.20	22.2	26.5	0.6	7.3	7.9
9C	11/7/69	8.7	5.0	9.50	2.00	-	-	2.5	12.4	14.9
9D	11/6/69	9.3	5.0	10.75	2.75	23.4	31.4	1.9	16.4	18.3
9E	11/5/69	8.9	5.0	9.00	3.50	17.2	23.6	0.7	23.9	24.6

and five responses. This study eliminated many variables and indicated that an upset condition occurred in the pulp mill during four runs. These runs were:

Run #	Date
10	12/10/69
16	12/11/69
1	12/12/69
14	12/19/69

Consequently, a second study was made using seventeen runs by eliminating these four runs. These two studies then allowed some contrasts to be made on the ability of the system to handle both "normal" feeds and "abnormal" feeds.

Table 23 page 83 shows the means and standard deviations of all the variables included in the twenty-one run study. Note that each parameter was studied over a wide range of values and that the responses also varied widely. This wide range illustrates the efforts that were made in the study to thoroughly examine the carbonation process system.

The simple correlation coefficient is a dimensionless measure of the amount of correlation between two variables. A zero value means no correlation. Direct correlation is indicated by a plus one, while indirect correlation is a minus one.

The correlation coefficient of the various independent variables with the dependent response ranks their importance when used as single variables to predict the response. The correlation coefficient for each pair of independent variables measures the degree of redundancy existing among the independent variables which can influence the study.

The significant correlations in this study were examined. Table 24 page 84 shows a list of variables that were eliminated from subsequent studies due to their high correlation with the primary design and operating parameters.

Table 25 page 85 shows the means and standard deviations of the remaining eleven independent variables and parameters and the five performance responses for all twenty-one runs.

TABLE 23

MEANS AND STANDARD DEVIATIONS OF ALL VARIABLES

Variable	Mean	Deviation	High	Low	Ratio	DEV/XBAR
FD-PH	12.1381	0.0920	12.3000	11.9000	0.96	0.
FD-TP	95.1438	11.9181	106.0024	63.0002	0.59	12.
COLOR	172.6212	43.1166	300.0022	120.0024	0.40	24.
FD-DI	7.9163	1.6552	9.2379	1.0149	0.10	20.
FD-SU	0.1452	0.1455	0.5590	0.0330	0.05	100.
FD-TO	8.0549	1.6821	9.2879	1.0209	0.10	20.
PH	9.3476	0.6925	10.4000	8.7000	0.83	7.
DEPTH	6.9000	2.4556	10.5000	5.0000	0.47	35.
RATE	13.4071	4.9139	20.9999	8.3000	0.39	36.
CYCLE	1.6666	1.6082	4.5000	0.0000	0.00	96.
STIR	60.7620	8.9325	64.0002	33.0002	0.51	14.
FLOW	13.8952	13.7485	45.0000	0.0000	0.00	98.
SOLID	408.0029	970.5128	3600.0214	0.0000	0.00	237.
CO2	21.8380	8.5189	50.5000	14.3000	0.28	39.
TEMPC	175.0969	83.8884	350.0021	50.0002	0.14	47.
TIME	19.2714	7.8743	38.8000	8.5000	0.21	40.
FLOC	9.1761	3.1429	14.1000	4.9000	0.34	34.
RISE	1.7157	0.6274	2.6899	1.0600	0.39	36.
TEMPD	90.3814	12.5839	104.0024	62.0002	0.59	13.
DE-DI	0.3377	0.1796	0.6340	0.0500	0.07	53.
DE-SU	1.2203	0.7527	3.1039	0.1170	0.03	61.
DE-TO	1.5581	0.7752	3.4049	0.6090	0.17	49.
SP.GR	1.0828	0.0434	1.1649	1.0109	0.86	4.
DAYS	24.2382	18.7960	63.0002	0.0000	0.00	77.
CO-CA	25.2380	7.1617	42.3000	16.3000	0.38	28.
CO-XX	5.4380	2.5218	10.5000	1.5000	0.14	46.
CO-TO	30.6761	8.2583	47.8000	20.8000	0.43	26.
CA-DI	3.8761	1.9929	6.9000	0.6000	0.08	51.
CA-SU	14.5857	9.2957	38.0000	1.3000	0.03	63.
CA-TO	18.4619	9.6376	41.6999	5.8000	0.13	52.

TABLE 24

LIST OF VARIABLES ELIMINATED DUE TO CORRELATION

FEED STREAM CHARACTERISTICS

	Code	
1.	FD-PH	Feed stream pH, dimensionless.
2.	FD-TR	Feed stream temperature, in degrees Fahrenheit.
3.	FD-DI	Feed stream dissolved calcium as $CaCO_3$, in pounds per 1000 gallons.
4.	FD-SU	Feed stream suspended calcium as $CaCO_3$, in pounds per 1000 gallons.

CARBONATOR OPERATING VARIABLES

5.	STIR	Carbonator agitator speed, in revolutions per minute.
6.	FLOW	Densator recycle flow, as percent of feed.
7.	SOLID	Densator recycle solids, as percent of feed.

DENSATOR OPERATING VARIABLES

8.	*	Densator agitator speed, in revolutions per minute.
9.	FLOC	Densator floculation time, in minutes.
10.	RISE	Densator rise rate, in gallons per square per minute.
11.	TEMPD	Densator temperature, in degrees Fahrenheit.
12.	DE-DI	Densator overflow dissolved calcium as $CaCO_3$, in pounds per 1000 gallons.
13.	DE-SU	Densator overflow suspended calcium as $CaCO_3$, in pounds per 1000 gallons.
14.	DE-TO	Densator overflow total calcium as $CaCO_3$, in pounds per 1000 gallons.

* Held constant.

TABLE 25

MEANS AND STANDARD DEVIATIONS OF SELECTED VARIABLES

Code	Mean	Deviation	High	Low	Ratio	Dev/XBAR
COLOR	172.6211	43.1166	300.0022	120.0023	0.40	24.
FD-TO	8.0549	1.6821	9.2879	1.0209	0.10	20.
PH	9.3476	0.6925	10.4000	8.7000	0.83	7.
DEPTH	6.9000	2.4556	10.5000	5.0000	0.47	35.
RATE	13.4071	4.9139	20.9999	8.3000	0.39	36.
CYCLE	1.6666	1.6032	4.5000	0.0000	0.00	96.
CO2	21.9285	8.5085	50.5000	14.3000	0.28	38.
TEMPC	179.8589	82.7910	350.0021	50.0002	0.14	46.
TIMEC	19.2714	7.8743	38.8000	8.5000	0.21	40.
SP.GR.	1.0843	0.0428	1.1649	1.0109	0.86	3.
DAYS	24.2382	18.7960	63.0002	0.0000	0.00	77.
CO-CA	25.3714	7.1002	42.3000	16.3000	0.38	27.
CO-TO	30.6523	8.2716	47.8000	20.8000	0.43	26.
CA-DI	3.8761	1.9929	6.9000	0.6000	0.08	51.
CA-SU	14.5857	9.2957	38.0000	1.3000	0.03	63.
CA-TO	18.4619	9.6376	41.6999	5.8000	0.13	52.

Table 26, page 87, shows the simple correlation coefficients between the variables. Note that nearly all these coefficients are close to zero, indicating independence (orthogonality) of the variables. Note also that some responses are correlated with other responses, meaning that they characterize similar things. This redundancy is illustrated in Figures 12 and 13 pages 88 and 89 and indicates that the system's performance can be characterized by Total CO_2 Transferred (reaction rate) and Total Calcium loss (efficiency).

Mathematical Models of System

Heuristic Regression Analysis was used to obtain the mathematical models of the dependent responses. A residual study was included for each model, questionable observations eliminated, and additional models developed. The previously mentioned four upset runs were eliminated and models developed. A study of data indicates that even though the individual data were variable, good and similar mathematical models were developed for the responses characterizing the carbonation pilot plant's performances during each run of both the twenty-one run study and the seventeen run study. It is beyond the scope of this report to explain the many statistical and computer concepts used. Only the conclusions are discussed in the subsequent sections and are primarily based on models using all the data.

A study of the resulting mathematical models indicates that several variables thought to be important actually had little or no effect on the system. Neither carbonation feed rate or carbonator retention time had an influence on the system's performance --- an indication that the pilot plant was not run at full design rate. Feed stream total calcium and gas feed CO_2 concentration had little or no effect on the system.

Reaction Rate Model

One of the major objectives of this pilot plant experimental program was to establish the best design conditions for promoting the reaction rate of CO_2 transferred to $CaCO_3$. A simple mathematical model was developed based on just two variables. Figure 14, page 90, shows a direct reading response plot of the effect of carbonation depth and Densator recycle rate on reaction rate measured as Total CO_2 transferred. A high rate of 46 percent can be obtained at 10.5 feet carbonation depth and a

TABLE 26

SIMPLE CORRELATION COEFFICIENTS

Code	Color	FD-TO	PH	Depth	Rate	Cycle	CO2	TEMPC	TIMEC	SP.GR
COLOR	1.00	0.18	0.10	-0.22	-0.10	0.13	0.01	0.14	-0.02	-0.21
FD-TO	0.00	1.00	0.06	0.14	-0.15	0.11	0.02	-0.20	0.20	-0.23
PH	0.00	0.00	1.00	0.12	0.27	0.02	0.46	-0.15	-0.03	0.16
DEPTH	0.00	0.00	0.00	1.00	0.22	0.06	-0.12	0.13	0.57	0.22
RATE	0.00	0.00	0.00	0.00	1.00	0.05	0.27	0.28	-0.50	-0.01
CYCLE	0.00	0.00	0.00	0.00	0.00	1.00	-0.38	-0.04	-0.31	-0.38
CO2	0.00	0.00	0.00	0.00	0.00	0.00	1.00	-0.40	-0.19	0.38
TEMPC	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.10	-0.33
TIMEC	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00	0.36
SP.GR	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.00
DAYS	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO-CA	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CO-TO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CA-DI	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CA-SU	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
CA-TO	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00

Days	CO-CA	CO-TO	CA-DI	CA-SU	CA-TO
-0.05	-0.12	-0.19	-0.01	0.43	0.41
-0.35	0.10	0.05	0.00	0.17	0.17
-0.03	0.21	0.12	0.40	0.43	0.49
0.38	0.55	0.66	0.42	-0.04	0.04
0.43	0.20	0.17	0.47	0.00	0.10
-0.29	0.59	0.57	-0.45	0.38	0.27
-0.18	-0.30	-0.30	0.34	0.13	0.19
0.37	-0.06	-0.03	0.03	-0.38	-0.36
0.08	0.00	0.09	0.21	-0.14	-0.09
0.09	-0.18	-0.17	0.59	-0.00	0.12
1.00	0.30	0.28	0.59	-0.20	-0.07
0.00	1.00	0.95	0.09	0.13	9.15
0.00	0.00	1.00	0.03	0.08	0.09
0.00	0.00	0.00	1.00	0.06	0.27
0.00	0.00	0.00	0.00	1.00	0.97
0.00	0.00	0.00	0.00	0.00	1.00

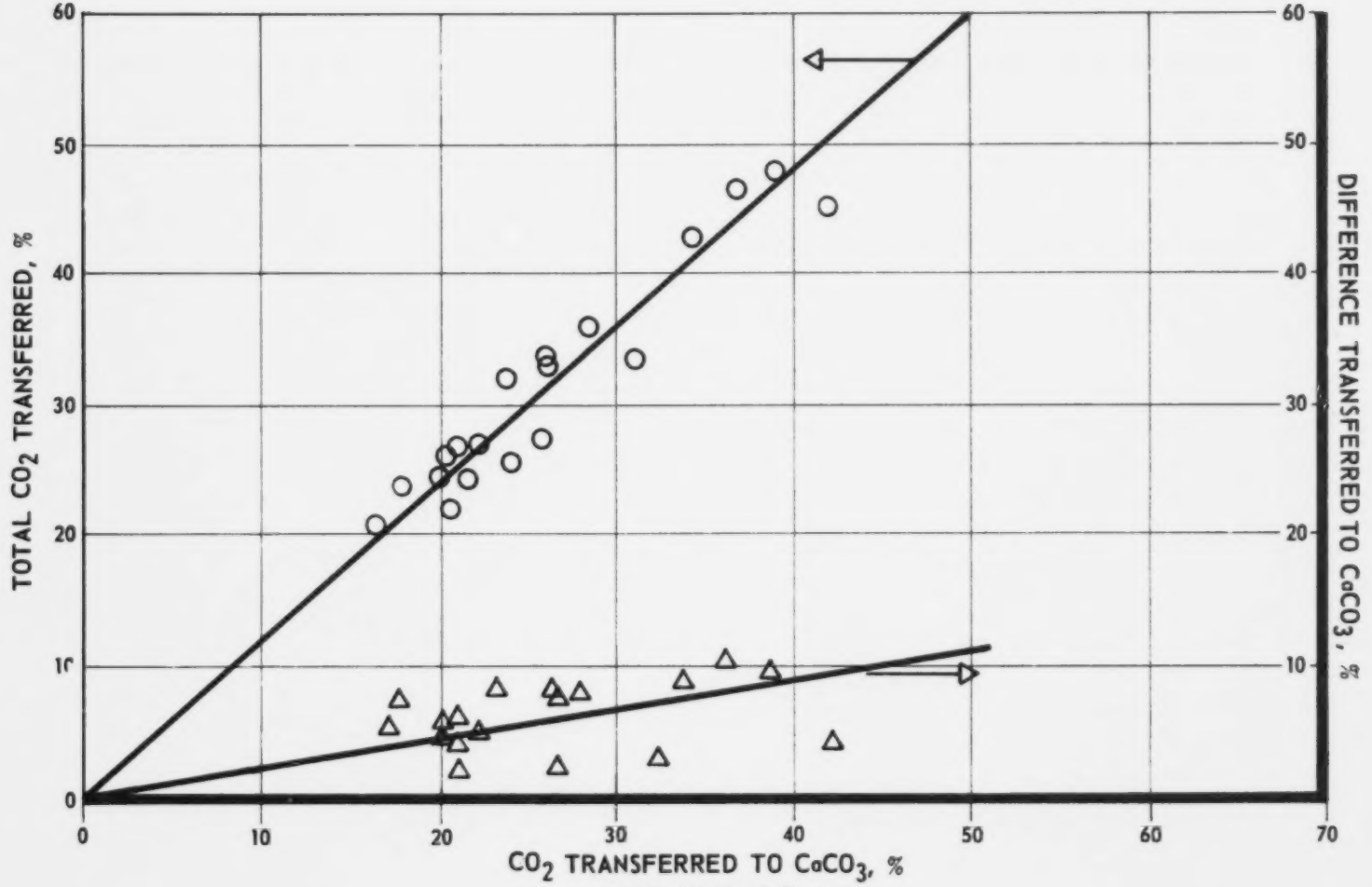


FIGURE 12
REACTION RATES

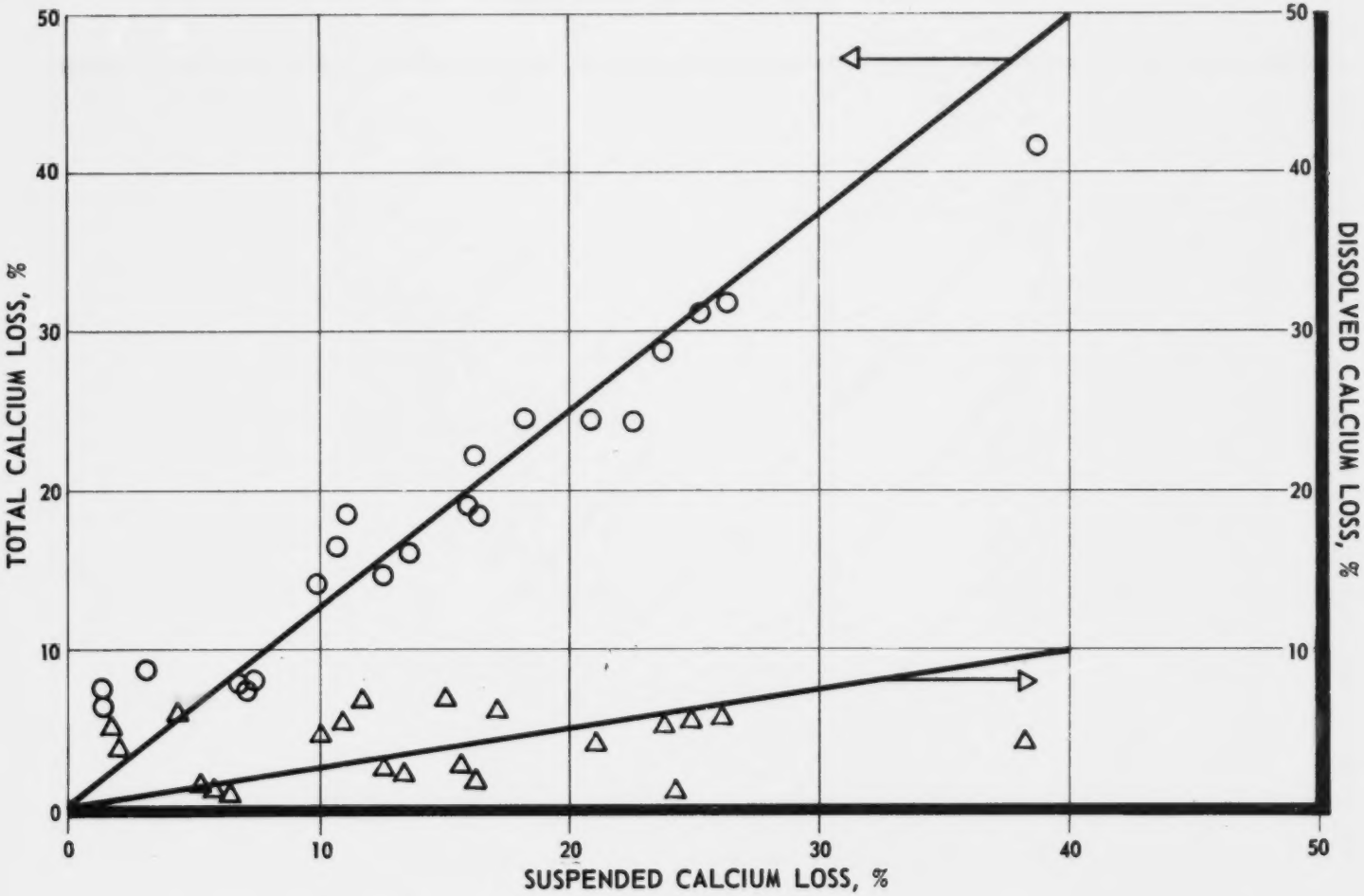


FIGURE 13
EFFICIENCY

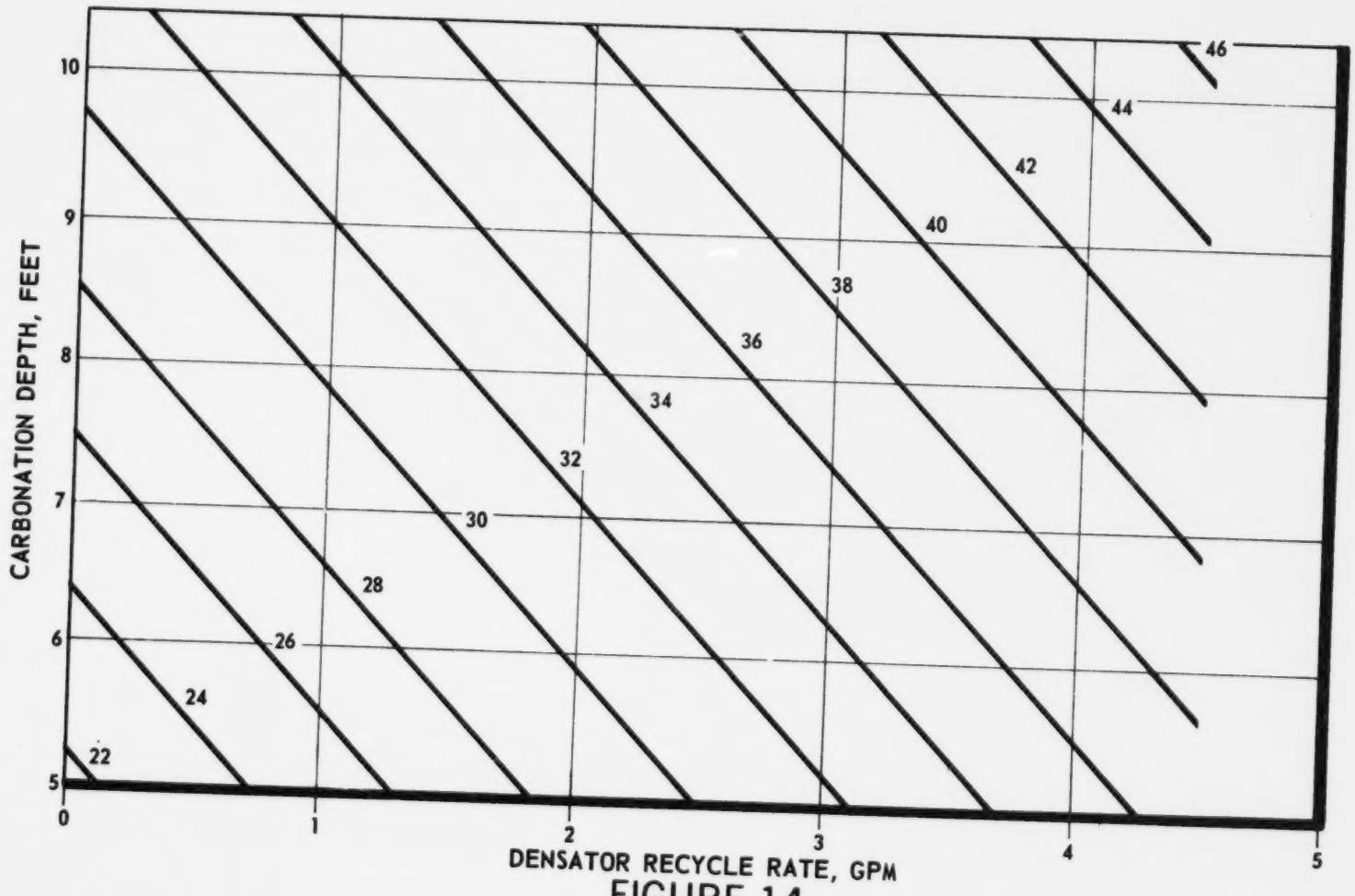


FIGURE 14

REACTION RATE (TOTAL CO_2 TRANSFERRED)

Densator recycle rate of 4.5 gpm. A more economical design from a capital investment view point would be 9.0 feet carbonation depth at a densator recycle rate of 6.5 gpm. This condition would give about 50 percent transfer.

System Efficiency Model

The carbonation-densator pilot plant efficiency and operating economics were influenced by three variables. High feed stream color, especially during the four runs in December 1969, when the pulp mill was upset, caused high calcium losses. Figure 15, page 92, shows a direct reading response plot of the effect of carbonation gas feed temperature and carbonation pH on the System Efficiency as measured by Total Calcium Loss. A low loss (high efficiency) of 5 percent can be obtained at 8.7 carbonation pH and a carbonation gas feed temperature of 325°F. At a lower gas temperature of about 200° F., the loss would rise to about 13 - 14 percent.

D. Carbonate Slurry

During the pilot plant shakedown period, it was found that the optimum setting for the Densator variable speed drive was for a shaft rpm of 10. This gave a peripheral speed on the thickener rake of 200 ft per minute. Agitation in the flocculation zone was also adequate at this speed to prevent settling in this area.

Blow-down times were set to give an underflow in the order of 0.33 percent of feed. This gave a slurry in the order of 20 percent solids or 2 lbs per gallon, and held a slurry depth of approximately 18 inches in the Densator.

The above settings were somewhat critical in that with higher settings the slurry bed would rise above the Densator down draft tube, carryover would increase and slurry underflow concentrations would decrease. With lower settings, the slurry blowdown would be too highly concentrated to flow by gravity.

Filtration test on the carbonate slurry gave filtration rates comparable to lime mud in the mill causticizing system. Figure 16, page 93, compares filtration rates vs percent solids. The carbonate solids have a distinct tannish color. However, the carbonate purity averaged 99.49 percent calcium carbonate on two samples collected during different weeks.

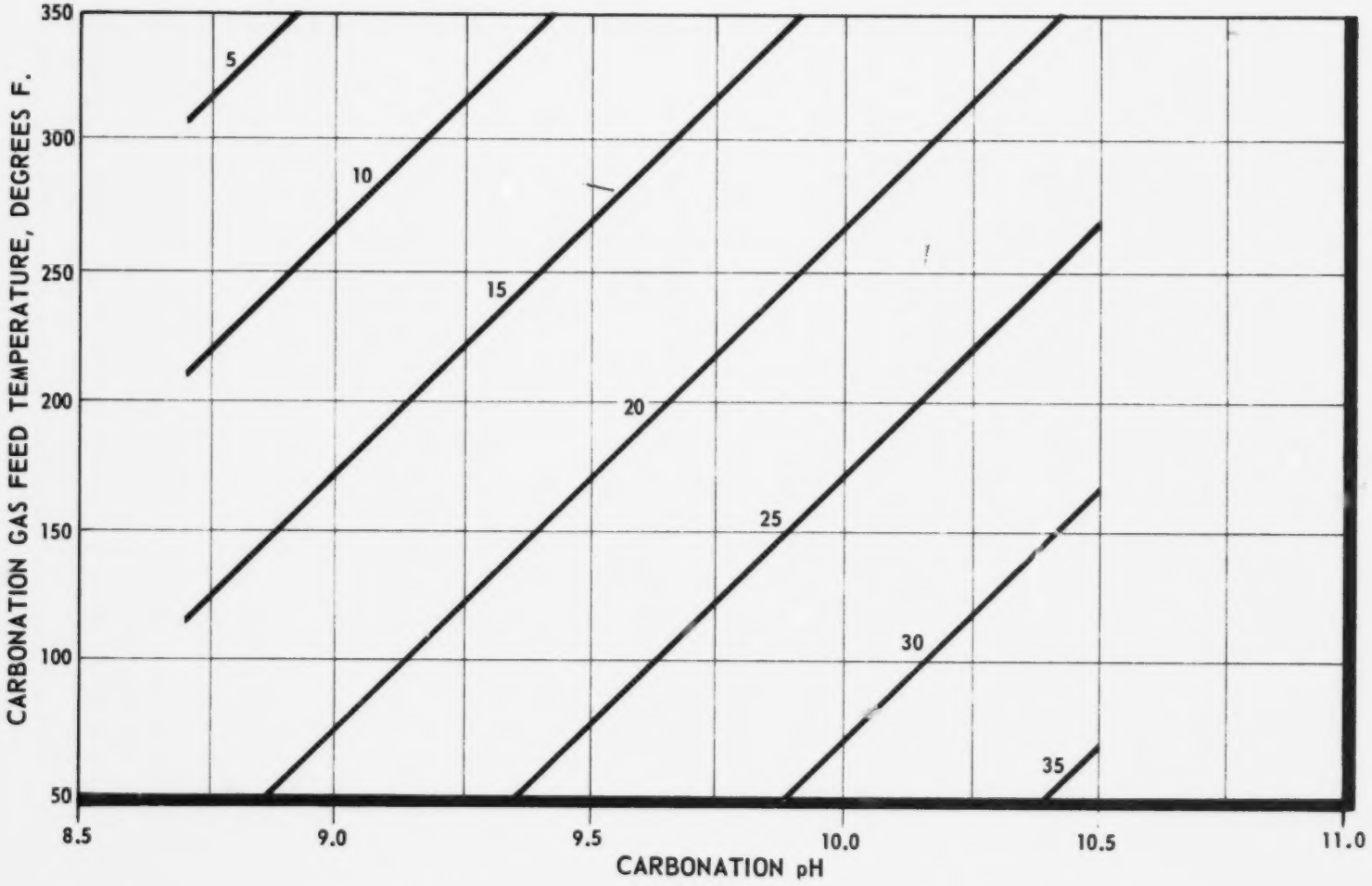


FIGURE 15
SYSTEM EFFICIENCY (TOTAL CALCIUM LOSS)

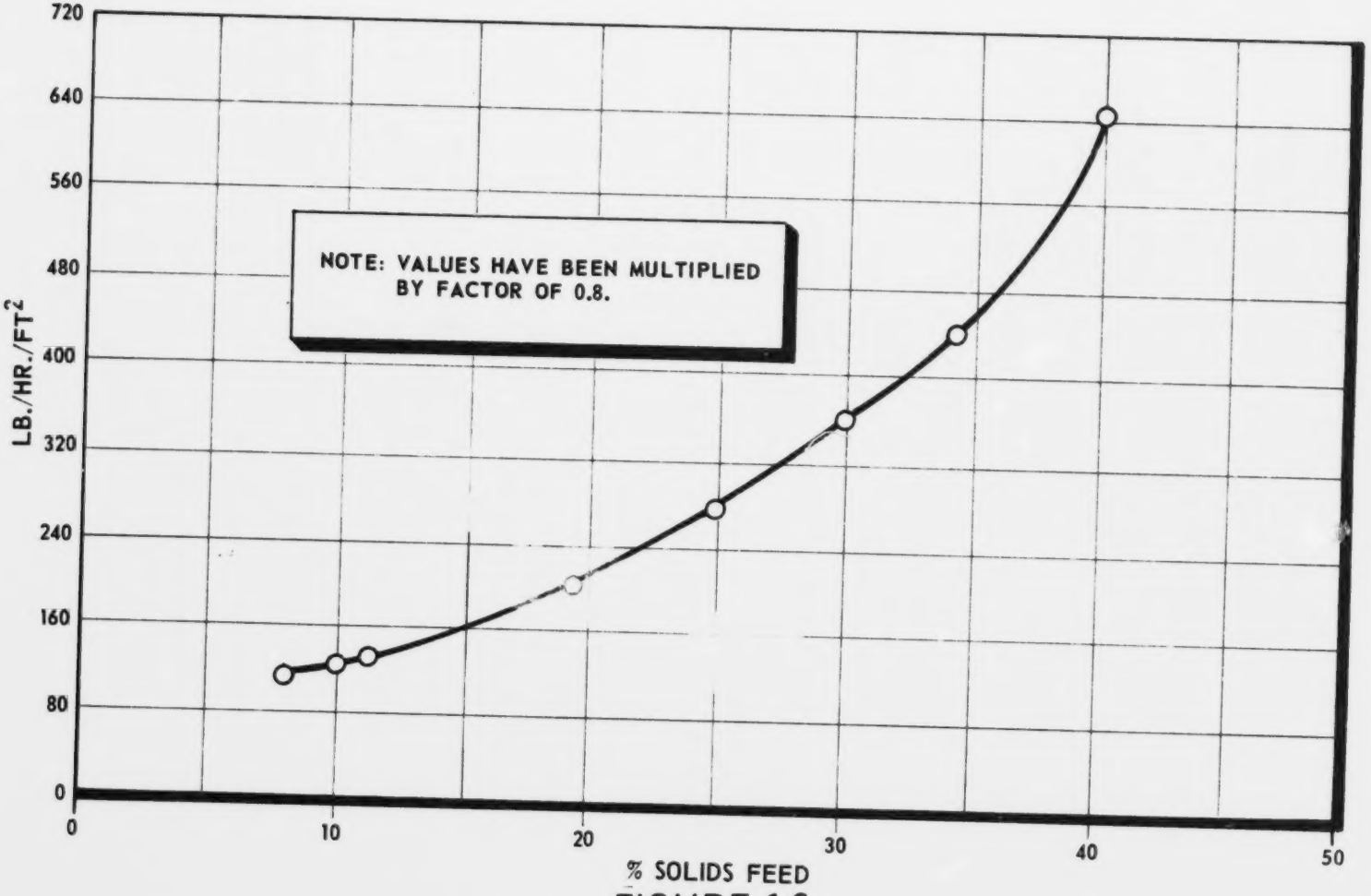


FIGURE 16
EIMCO FILTER LEAF TESTS DENSATOR SLUDGE

Based on the filterability and purity, there would be no problem of calcining the carbonator solids in the mill lime kiln.

E. Instrumentation Evaluation

The pilot plant instrumentation performed satisfactorily after adjustment and final tune-ups with exception of the turbidity measurement device. In this one exception, the instruments were well designed. The unsatisfactory performance was due to misapplication. If the horizontal flow through reference cell could be replaced with one of perpendicular flow through design, the problem of calcium carbonate disposition would be eliminated. The instrument could then be used in controlling the carbonate slurry level in the clarification zone of the system as well record the quality of the treated waste.

With the above modification, all of the instruments and controls can be used with excellent results in a full scale carbonation plant.

F. Summarization of Carbonation Process

The pilot plant adequately served its purpose in that it made it possible to study the many variables of the carbonation process under actual mill operating conditions. Flows were such that scale up to handle the total mill waste can be made on an economical design to meet the full range of mill demands.

The operating problems and modifications experienced in the startup were realistic in that they simulated what could have been experienced with a full scale plant without the benefit of the prior pilot runs.

There was a close agreement with other investigators of the carbonation process. However, the computer analysis emphasized the importance of keeping the formation of colloidal calcium carbonate to a minimum. The five variables and efficiency all pertained to the physical and chemical reaction process of solid calcium formation. Variables such as carbonation retention time, Densator flocculation time and rise rates which were thought to be of major consideration were found to have no influence on the process within the operating range of the pilot plant.

The calcium carbonate product is readily concentrated and has filtration characteristics comparable to mill causticizing mud. As it has a high purity, there are not foreseeable problems in its reuse in the mill system.

Some "do's" and "don't's" which are based on operating experience are:

1. Maintain calcium carbonate solids concentration in the carbonation zone in the range of five to six thousand parts per million pounds of total treated effluent. This may be conservative as this was the solids concentration at which the ratio of solids to unreacted gases gave a minimum foaming tendency in the pilot plant test runs.

2. Do not use orifice type diffusers as openings up to 3/8 inch will plug within a few days. There may also be problems with turbine type diffusers as calcium deposits could cause imbalance of the turbine blades. The most practical type of diffusers appears to be of a slotted design similar to the one used in the pilot plant. In any case provisions should be made for cleaning and servicing without having to shut the plant down.

3. Instrumentation for turbidity measurement should have a perpendicular flow through reference cell with provisions for intermittent cleaning.

4. With recycle to the carbonator, a combined carbonation and clarification unit would have a definite advantage over separate units in that one mechanical rake for solids removal would serve both steps in the process.

SECTION VII

ACKNOWLEDGMENT

The color removal system was developed and designed as an integral part of the tertiary treatment process for kraft mill effluent by Mr. W. J. Verross, Vice President and General Manager of Interstate Paper Corporation and Mr. J. G. Vamvakias of The Rust Engineering Company, Pittsburgh, Pa. It was through their efforts and the cooperation of Mr. R. S. Howard, Director of the Georgia State Water Quality Control Board, and Mr. Charles H. Starling, Director of Industrial Waste Service, that the treatment system became a reality.

The assistance of Mr. E. P. Lomasney, Project Cfficer of the Federal Water Quality Administration, Atlanta, Georgia, and the guidance given by Mr. George R. Webster, Federal Water Quality Administration, Washington, D. C., has contributed materially to the successful operation of the color removal system.

The very necessary environmental guidelines have been provided by Mr. H. B. Counts of the U. S. Geological Survey and his staff and Mr. M. D. Dahlberg and co-workers of the University of Georgia Marine Institute. Their findings in their follow up investigations have provided a much needed encouragement.

The carbonation pilot plant study was a joint project of Interstate Paper Corporation and Continental Can Company. Mr. R. L. Scoville, Manager Special Projects, Continental Can Company, provided the necessary manpower from his technical staff and contributed materially in his close association with the project.

Mr. Floyd A. Miller, of Glidden-Durkee Division of SCM Corporation, assisted in setting up the statistically designed program and provided the computer analyses of the operating variables and responses monitored during the program. His report of the computer analyses is incorporated in this report as submitted.

The program was under the direction and supervision of Charles L. Davis, Jr., Pollution Control Director, Interstate Paper Corporation. The responsibility for the program has been with Mr. Edgar L. Hart, Jr., Utilities Manager, Interstate Paper Corporation. His support and administrative assistance provided the necessary team effort.

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SECTION IX

APPENDIX

TABLE I
COLOR REDUCTION VS. LIME FEED
JUNE 27, 1968 - SEPTEMBER 30, 1968

Untreated Waste		Average Color Reduction (ppm) per 100 ppm $Ca(OH)_2$							
Color Range		200-500	600-1000	1100-1500	1600-2000	2100-2500	3,000	7,000	8,000
Av. Color		423	795	1320	1800	———	3,000	———	———
Lime Feed ppm $Ca(OH)_2$	1400	—	(1) 49.8	—	—	—	—		
	1500	—	(2) 54.7	—	—	—	(1) 192.7		
	1600	—	(7) 41.1	(2) 79.8	—	—			
	1700	(1) 15.3	(6) 39.1	(3) 68.3	(1) 111.3	—			
	1800	(1) 7.4	(16) 39.2	(1) 61.3	(1) 81.5	—			
	1900	(5) 20.0	(10) 39.6	(3) 69.7	—	—	(1) 151.4		
	2000	(1) 17.4	(3) 31.4	—	(1) 83.2	—	—		(1) 396.0
	2100	(1) 21.4	(3) 39.0	—	—	—	—	(1) 311.9	
	2200	(1) 18.9	(2) 30.8	(1) 44.4	—	—	—		
	2300	(1) 19.3	(1) 21.8	—	(1) 72.0	—			
	2400	—		—	—	—	—		
	2500	—		—		—	—		
Treated color									
Av		(11) 72	(51) 86	(10) 118	(4) 131		(2) 108	(1) 450	(1) 80
Max.		150	125	200	150		125	450	80
Min.		40	50	60	100		90	450	80

(-) Number of days.

TABLE 2
COLOR REDUCTION VS. LIME CONCENTRATION
OCTOBER 1, 1968 - OCTOBER 31, 1968

Untreated Waste		Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		200-500 400	600-1000 796	1100-1500 1225	1600-2000 2000	2100-2500	3000	4000	5000
Lime Concentration ppm $Ca(OH)_2$	1400	—	—			—			
	1500	—	—			—			
	1600	—	—			—			
	1700	—	(2) 35.6			—			
	1800	—	(5) 40.9	(1) 62.8		—			
	1900	—	—	—		—			
	2000	(1) 17.6	(2) 35.0	—		—			(1) 232.0
	2100	—	(2) 29.7	(1) 53.4		—			
	2200	—	—	—		—			
	2300	—	(1) 41.6	—		—	(1) 121.5		
	2400	—	—	(2) 46.6		—			
	2500	—	(1) 29.1		(1) 75.4	—		(1) 145.3	
	2600		(1) 29.6	—	—	—	(1) 106.6		
Treated Color Av.		(1) 50	(14) 87	(4) 114	(1) 150	—	(2) 200	(1) 300	(1) 300
Max.		50	200	150	150	—	250	300	300
Min.		50	50	75	150	—	150	300	300

TABLE 3

COLOR REDUCTION VS. LIME CONCENTRATION
November 1, 1968 - November 30, 1968

Untreated Waste		Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		200-500 388	600-1000 764	1200	1500	3000	6000		
Lime Concentration ppm $Ca(OH)_2$	1200	(1) 32.0	(1) 57.6						
	1300		(5) 52.1			(1) 215.4			
	1400		(3) 46.5						
	1500	(2) 22.1	(5) 41.3		(3) 93.1				
	1600		(5) 41.8	(1) 65.7			(1) 372.8		
	1700		(1) 40.8						
	1800		(1) 32.8						
Treated Color		(3)	(21)	(1)	(3)	(1)	(1)		
Av.		77	104	150	108	200	125		
Max.		90	200	150	150	200	125		
Min.		70	50	150	75	200	125		

TABLE 4
COLOR REDUCTION VS. LIME CONCENTRATION
DECEMBER 1, 1968 - DECEMBER 31, 1968

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm Ca $(OH)_2$							
Color Range Av. Color		200-500 500	600-1000 818	1100-1500 1261	1600-2000 1890	2750 2750			
Lime Concentration ppm Ca $(OH)_2$	1100		(1) 80.0						
	1200								
	1300	(1) 32.9	(2) 45.1	(1) 78.4		(1) 197.7			
	1400		(2) 42.8	(2) 69.1	(1) 72.5				
	1500		(4) 44.2						
	1600			(2) 68.0					
	1700								
	1800								
	1900		(1) 33.6	(1) 66.8	(2) 91.5				
	2000			(2) 61.1	(2) 77.5				
	2100		(1) 42.8		(1) 71.7				
Treated Color Av.		(1) 75	(11) 135	(8) 158	(6) 178	(1) 200			
Max.			250	250	250	200			
Min.			75	80	90	200			

( ) - Number of days

TABLE 5
COLOR REDUCTION VS. LIME CONCENTRATION
JANUARY 1, 1969 - JANUARY 31, 1969

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm $Ca(OH)_2$							
Color Range Av. Color		600-1000 820	600-1000 880	1100-1500 1300	1600-2000 1890	2100-2500 2270	2860	3600	
Lime Concentration ppm $Ca(OH)_2$	800		(1) 58.6						
	1100		(2) 66.5						
	1200								
	1300	(2) 57.2					(1) 165.1	(1) 273.8	
	1400	(2) 50.5	(1) 56.8						
	1500		(1) 53.5			(1) 159.3			
	1600				(1) 112.9				
	1700	(1) 42.1		(1) 82.1					
	1800			(3) 58.5	(1) 85.9	(1) 107.9			
	1900		(2) 47.3	(1) 56.1					
	2000		(1) 40.4						
			Modified Lime Feed System →						
Treated Color Av. Max. Min.		(5) 94 125 75	(8) 127 190 100	(5) 138 220 90	(2) 188 200 175	(2) 155 180 130	(1) 200	(1) 150	

( ) Number of days

TABLE 6
COLOR REDUCTION VS. LIME CONCENTRATION
FEBRUARY 1, 1969 - FEBRUARY 28, 1969

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm $Ca(OH)_2$							
Color Range Av. Color		600-1000 920	1100-1500 1275	1600-2000 1914	2100-2500 2313	5000	6800		
Lime Concentration ppm $Ca(OH)_2$	1000		(1) 104.4	(1) 176.0	(1) 238.8				
	1100	(2) 66.7		(1) 113.5					
	1200			(1) 118.8	(1) 177.4				
	1300	(1) 51.4		(1) 133.8					
	1400								
	1500		(1) 72.7	(2) 115.1	(1) 141.3				
	1600		(1) 62.6						
	1700								
	1800						(1) 389.2		
	1900	(1) 47.3							
	2000	(1) 46.6							
	2100				(1) 104.4				
	2200					(1) 227.1			
	2300		(1) 60.7						
	2400								
	2500								
	2600								
	2700			(1) 70.8					
Av.		(5) 141	(4) 143	(7) 191	(4) 193	(1) 200	(1) 125		
Max.		200	180	300	250	200	125		
Min.		80	90	100	170	200	125		

( ) Number of days.

TABLE 7

COLOR REDUCTION VS. LIME CONCENTRATION
March 1, 1969 - March 31, 1969

Untreated Waste		Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		200-600 410	600-1000 900	1000-1500 1219	1500-2000 1689	2000	2640	3200	
Lime Concentration ppm $Ca(OH)_2$	500			(1) 266.6					
	600								
	700								
	800		(1) 103.1	(1) 138.7					
	900								
	1000			(1) 84.2					
	1100					(1) 142.8	(1) 199.3		
	1200								
	1300		(1) 67.7	(2) 94.5					
	1400	(1) 13.2	(2) 49.4						
	1500	(1) 15.8		(2) 62.2					
	1600	(1) 26.8	(2) 53.9	(1) 75.7	(1) 110.7				
	1700			(2) 60.9				(1) 173.7	
	1800	(2) 22.4			(1) 73.1				
	1900	(1) 16.0			(1) 81.0				
	2000		(2) 43.7						
		(6)	(8)	(10)	(3)	(1)	(1)	(1)	
Av.		83	90	123	87	225	240	300	
Max.		100	140	220	130	225	240	300	
Min.		60	20	50	30	225	240	300	

( ) Number of days.

TABLE 8

COLOR REDUCTION VS. LIME CONCENTRATION

APRIL 1, 1969 - APRIL 30, 1969

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm $Ca(OH)_2$							
Color Range Av. Color		200-600 389	600-1000 839	1000-1600 1344	1600-2000 1707	2280 2280	4200 4200		
Lime Concentration ppm $Ca(OH)_2$	800		(1) 109.3						
	900								
	1000			(1) 123.7					
	1100		(1) 63.8	(1) 114.4					
	1200	(2) 19.9		(1) 107.0					
	1300	(2) 33.8	(3) 66.3	(2) 87.0			(1) 311.5		
	1400	(1) 23.0	(2) 52.4		(1) 108.1				
	1500	(1) 17.7	(1) 51.8	(2) 86.1	(1) 99.9				
	1600	(1) 7.0	(2) 39.0		(1) 104.6	1) 134.0			
	1700	(1) 30.1							
Treated Waste Color									
Av.		(8) 65	(10) 81	(7) 123	(3) 160	(1) 80	(1) 160		
Max.		140	120	180	200	80	160		
Min.		30	40	60	100	80	160		

( ) - No. Days

TABLE 9
COLOR REDUCTION VS. LIME CONCENTRATION
May 1, 1969 - May 31, 1969

Untreated Waste	Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color	200	200-600	600-1000	1000-1400				
900			(2) 68.8					
1000			(3) 63.4					
1100			(1) 75.9					
1200		(1) 23.0	(1) 75.8	(2) 87.2				
1300	(1) 6.2		(3) 47.3	(2) 75.3				
1400	(2) 8.4	(2) 26.8	(2) 40.8	(2) 75.2				
1500			(2) 49.1					
1600								
1700		(1) 18.2		(1) 60.3				
1800								
1900								
2000			(2) 29.0					
2100								
Treated Waste	(3)	(4) 75	(16)	(7)				
Color Av.	50	100	92	96				
Max.	60	50	150	120				
Min.	40		50	60				

( ) number of days.

TABLE 1^
COLOR REDUCTION VS LIME CONCENTRATION
JUNE 1, 1969 - JUNE 30, 1969

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm $Ca(OH)_2$							
Color Range Av. Color		200-600 489	600-1000 937	1000-140(1180	1400-1800 1607	2400-2500 2450	3000-3200 3067		
Lime Concentration ppm $Ca(OH)_2$	1000	(2) 41.3	(1) 78.9						
	1100			(1) 91.6			(1) 262.6		
	1200		(1) 73.7		(2) 117.2		(1) 232.8		
	1300	(1) 32.4	(2) 78.2						
	1400	(1) 35.7	(2) 65.1	(4) 73.3		(2) 169.4			
	1500	(2) 22.8							
	1600	(2) 21.5	(1) 51.0	(1) 76.1					
	1700								
	1800	(1) 25.4			(1) 123.7		(1) 166.8		
Treated Waste Color % Red. Av. Max. Min.		81.5 (9) 91 100 65	90.5 89 110 60	90.2 (6) 115 150 90	94.4 (3) 90 100 80	93.5 (2) 160 170 160	94.2 (3) 175 200 150		

(-) No. days

TABLE 11
COLOR REDUCTION VS LIME CONCENTRATION
JULY 1, 1969 - JULY 31, 1969

Untreated Waste		AVERAGE COLOR REDUCTION (ppm) PER 100 ppm $Ca(OH)_2$							
Color range Av. Color		600-1000 853	1000-1400 1146	1400-1800 1603	2200	6000			
Lime Concentration ppm $Ca(OH)_2$	700	(1) 84.7		(1) 207.8					
	800		(1) 127.5	(1) 180.4					
	900								
	1000								
	1100		(2) 89.8	(1) 117.4					
	1200	(2) 59.8	(1) 99.2						
	1300	(2) 58.2		(1) 118.8		(1) 454.0			
	1400		(3) 76.0						
	1500	(2) 50.6	(1) 65.4		(1) 136.3				
	1600	(1) 54.7	(1) 54.7	(1) 84.5					
	1700			(1) 91.5					
	1800		(1) 54.4						
Treated Waste Color % Red. Av. Max. Min.		87.6 (8) 105 108 89	89.1 (10) 125 180 85	90.7 (6) 148 180 110	93.2 (1) 150 150 150	97.1 (1) 175 175 175			

(-) No. days

TABLE 12

COLOR REDUCTION VS. LIME CONCENTRATION

August 1, 1969 - August 31, 1969

Untreated Waste	Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color	200-600	600-1000 903	1000-1400 1209	1400-1800 1560	600-1800 1130			
Lime concentration ppm $Ca(OH)_2$ *800			(1) 138.4					
*900		(1) 91.6	(1) 146.2					
1000								
1100								
1200		(4) 72.7	(3) 96.5	(1) 128.0				
1300		(2) 60.8	(5) 74.0	(1) 109.5				
1400		(1) 55.4	(5) 80.2					
1500			(2) 68.5					
1600		(2) 42.9	(1) 84.3					
1700			(1) 57.8					
Treated Waste		(10)	(19)	(2)	(31)			
Color % Red. Av.		88.2	90.5	89.3	89.9			
Max.		106	114	175	115			
Min.		130	130	200	200			
		80	80	150	80			

(-) No. Days

* Diluted by surface run-off from heavy rains.

Note - Both untreated & treated waste colors reported at pH 7.6

TABLE 13

COLOR REDUCTION VS. LIME CONCENTRATION
September 1, 1969 - September 30, 1969

Untreated Waste		Average Color Reduction ppm per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		1000-1400 1216	1400-1800 1636	1800-2200 1993	1000-2200 1475				
Lime Concentration ppm $Ca(OH)_2$	800								
	900	(1) 135.5							
	1000		(1) 161.7						
	1100	(1) 76.4							
	1200	(1) 70.0	(6) 118.4						
	1300			(2) 145.5					
	1400	(2) 79.2	(1) 103.7						
	1500	(3) 81.0							
	1600	(2) 63.4							
	1700								
	1800								
	1900								
	2000	(1) 54.3							
	2100								
	2200								
	2300								
	2400								
	2500			(1) 71.4					
Treated Waste		(11)	(8)	(3)	(22)				
Color % Red.		89.3	91.1	91.8	90.4				
Av.		126	146	157	138				
Max.		190	180	240	240				
Min.		90	110	100	90				

(-) No. of days

Note - Untreated waste color unadjusted pH 12.3
Treated waste color adjusted pH 7.6

TABLE 14

COLOR REDUCTION VS. LIME CONCENTRATION
October 1, 1969 - October 31, 1969

Untreated Waste		Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		1000-1400 1217	1400-1800 1622	1800-2200 1962	3080 3080	1000-3080 1695			
Lime Concentration ppm $Ca(OH)_2$	800		(1) 203.7						
	900	(1) 108.2	(1) 179.6						
	1000			(1) 172.4					
	1100	(1) 83.9	(1) 139.4						
	1200	(1) 90.2	(3) 122.8	(1) 140.3					
	1300	(1) 93.4	(2) 105.7	(3) 142.4					
	1400	(1) 90.7		(2) 129.5					
	1500			(1) 124.1					
	1600	(1) 67.1							
	1700								
	1800		(1) 81.0						
	1900				(1) 153.9				
Treated Waste		(6)	(9)	(8)	(1)	(24)			
Color % Red.		89.8	91.4	92.1	95.1	91.4			
Av.		124	141	155	150	142			
Max.		150	200	180	150	200			
Min.		90	80	130	150	80			

(-) No. Days

Note - Untreated waste color unadjusted pH - 12.3
Treated waste color adjusted pH - 7.6

TABLE 15

COLOR REDUCTION VS. LIME CONCENTRATION
November 1, 1969 - November 30, 1969

Untreated Waste		Average Color Reduction (ppm) Per 100 ppm $Ca(OH)_2$							
Color Range Av. Color		1000-1400 1214	1400-1800 1534	1800-2200 2000	1000-2200 1520				
Lime Concentration ppm $Ca(OH)_2$	1000	(2) 112.1	(3) 147.2	(6) 159.0					
	1100	(2) 96.2	(3) 120.2						
	1200	(2) 87.9	(4) 113.9	(2) 140.5					
	1300	(4) 80.0	(1) 113.0	(1) 124.1					
	1400		(1) 93.8						
	1500								
	1600		(1) 91.0						
Treated Waste		(10)	(13)	(6)	(29)				
Color % Red.		87.2	90.4	90.2	89.3				
Av.		149	147	196	160				
Max.		200	190	240	240				
Min.		120	110	160	110				

(-) No days

Note untreated waste color at unadjusted pH 12.2
Treated waste color at adjusted pH 7.6

TABLE 16

COLOR REDUCTION VS. LIME CONCENTRATION

December 1, 1969 - December 31, 1969

Untreated Waste		Average Color Reduction (ppm) per 100 ppm $Ca(OH)_2$							
Color Range		1000-1400	1400-1800	1800-2200	2200-2600	3000	1000-3000		
Av. Color		1195	1766	2050	2467	3000	1849		
Lime Concentration ppm $Ca(OH)_2$	1000	(1) 124.3		(1) 189.9					
	1100		(1) 120.7	(1) 167.6					
	1200	(1) 89.8	(5) 135.6	(1) 148.1	(2) 179.4				
	1300	(2) 85.8			(1) 183.6				
	1400		(1) 116.4	(1) 151.8		(1) 188.3			
	1500		(1) 114.2						
	1600								
	1700								
	1800								
	1900								
	2000								
	2100								
	2200		(1) 75.9						
	2300	(1) 35.7							
Treated Waste Color % R_ed.		86.1	90.2	91.4	87.5	90.0	89.1		
Av.		(5) 170	(9) 174	(4) 175	(3) 253	(1) 300	(22) 190		
Max.		200	200	220	300	300	300		
Min.		130	110	160	200	300	110		

(-) No days

Note untreated waste color at unadjusted pH 12.2
Treated waste color at adjusted pH 7.6

*U.S. GOVERNMENT PRINTING OFFICE: 1972-484-485/96